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/THE GASIFICATION OF BIOMASS
IN COMMERCIAL DOWNDRAFT GASIFIERS/

by

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CHAPTER 1

INTRODUCTION

Increasing prices coupled with the instability and uncertainty in the supply of fossil fuels and diminishing reserves has prompted the search for alternate energy sources. In response, the conversion of biomass to energy has gained attention. Biomass is renewable and generally low in sulphur. Because it is renewable, it can be used without increasing the CO_2 content of the atmosphere, and the low sulphur content is an asset for small-scale utilization. The main disadvantages of biomass compared to fossil fuels are its wide distribution (non-point source) and low energy density. The non-point source nature of biomass makes large-scale processing costly due to high transportation costs. The low energy density requires larger quantities to be processed relative to fossil fuels for a given energy need. In this regard, beneficiation processes (see, e.g., Bain, 1980) are available for improving the properties of biomass. Wood pallets are a prime example of improved biomass.

Wood is a major source of biomass. Currently, wood provides about 2% of our total energy needs and could contribute up to 8% within the next decade (Zerba, 1981). The annual harvest of woody biomass amounts to about 1.4 billion tons in the US alone. Over 700 million tons of this material is not used because it is not of the right species, size, fiber length, fiber morphology etc. (Goldstein, 1978). Thus the potential for the utilization of wood to ensure a continuous supply of fuels and chemicals is significant.

Direct combustion of biomass is generally inconvenient and usually environmentally unacceptable. Hence various technologies have been investigated to convert biomass into more attractive fuels or chemicals. Among these, gasification technologies are of particular interest due to the versatility of gas and the convenience of use. Various types of reactors have been used to gasify biomass; they include fixed beds, moving beds, entrained beds, rotary kilns and fluidized beds. One of these technologies, the moving

bed downdraft gasifier, was studied in this thesis.

The objective of this thesis is to report on studies on the gasification of woody biomass in two similar commercial moving bed downdraft gasifiers. The study gathered complete material balance data and evaluated various performance measures for the gasifiers. These included various input and output stream rates, gas composition, material balance closure, gas yield, gas heating value, total energy output, mass conversion efficiency and cold gas efficiency. The influence of feed rate, type of feedstock, grata rotation speed and type of bed support on the gasifier performance was also investigated. Furthermore, an extensive mass and energy analysis was conducted based on the material balance data from one of the gasifiers.

A literature review is presented in Chapter 2. This review focussed on the gasification of biomass in moving bed downdraft gasifiers, including experimental studies and modeling efforts. Since only biomass of the macro-particle size is suitable for gasification in the downdraft gasifier, models for the pyrolysis of a single large biomass particle were reviewed, as a precursor to overall gasifier modeling. The single large particle models were classified into volumetric reaction and surface reaction models.

Chapter 3 presents an experimental study on the gasification of wood chips in the first gasifier. This study investigated the effect of feed rate on the gasifier performance. The feed rate was proportional to the fan speed and was varied from 27 to 126 kg/hr. Adequate data were gathered to evaluate material balance closures for the experiments. The data were also used to evaluate various gasifier performance measures.

Chapter 4 presents a detailed mass and energy analysis based on the material balance data obtained in Chapter 3. The mass analysis resulted in an

experimental stoichiometry for the gasification of wood chips. This was accomplished through a two level adjustment of the original data which forced both the overall and elemental balances to perfect closure. The energy analysis was based on the thermodynamic first and second laws. It evaluated the first law and second law efficiencies for the process for various operation modes (combinations of different usable products and output temperatures). Various energy dissipations were also determined, including the system heat loss, the exergy loss due to the heat loss and the exergy dissipation due to irreversibilities of the process.

Chapter 5 presents an experimental study on the gasification of wood materials in the second gasifier, similar to the one used in Chapter 3. In addition to the complete material balance data and evaluation of various performance measures, this work focussed on the influence of some operating parameters on the gasifier performance. The operating parameters studied include the type of feedstock, the grate rotation speed, and the type of bed support. Wood chips and wood pellets were used as feedstocks and the grate rotation speed was varied from 3 to 14 min/rev. Bed supports consisted of a 9 cm layer of ceramic balls (2 cm diameter) on top of the grid plate (grate), the grid plate with half of its available open area obstructed and the unobstructed grid plate.

Chapter 6 summarizes the major conclusions of this thesis and outlines the recommendations for extensions of this work.

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CHAPTER 2

LITERATURE REVIEW

INTRODUCTION

The increasing price and the uncertainty in the supply of fossil fuels in the 70's increased the need to search for alternative energy sources. In response to this, the utilization of all kinds of organic wastes, including biomass, has gained attention. Biomass has the potential of providing up to 20 fne1-quad equivalents per year in the US (Imman,1980); furthermore, it can also serve as a feedstock for industrial chemicals.

Biomass is renewable and, because of this, its utilization will not affect the composition of atmosphere, especially CO₂, which has been recognized as a potential threat to the entire planet. Furthermore, its very low sulphur content is an asset for small-scale utilization. On the other hand, its wide distribution (non-point source), low energy density and high degree of diversity (species) are the main disadvantages compared to fossil fuels. The non-point source nature of biomass makes large-scale processing costly due to high collection and transportation costs. Its low energy density requires larger quantities to be processed in order to obtain the same amount of energy. The wide variety of species further complicates the processing of biomass materials.

Direct combustion of biomass is generally inconvenient and usually environmentally unacceptable. Hence various technologies have been investigated to convert biomass into more attractive fuels or chemicals. Among these, gasification technologies, which convert biomass into low BTU gas, are of particular interest, due to the versatility of gas and the convenience of use. The superiority of gaseous fuels over solid fuels is evidenced by the fact that most existing industrial and residential heating equipment are designed for gas. Gasification of biomass is one way to

supplement our supply of gaseous fuels.

Gasification is a rather loosely defined term. Generally speaking, any process which converts organic materials (biomass as well as solid fossil fuels) into gas as well as minor amounts of liquid hydrocarbons and solid residues, may be called gasification. Gasification processes generally fall into two categories, thermochemical and biochemical. In this thesis, gasification will refer to thermochemical gasification which can be defined as the process by which solid organic materials are decomposed at elevated temperatures into gases as well as small amounts of liquid hydrocarbons and solid residues with the introduction of controlled amounts of oxidizing agents and heat. With the introduction of oxidizing agents and heat in various amounts, it bears different names: pyrolysis, with only heat introduced; pyrolysis-gasification, with both oxidizing agents and heat introduced; air or oxygen gasification, with air or oxygen introduced as the oxidizing agent; and in the extreme case, combustion, with enough oxidizing agent introduced to insure complete oxidation. Reed et al.(1980) have presented an extensive survey on the principles of biomass gasification, including the pyrolysis of biomass, the thermodynamics of gas-char reactions and the kinetics of char gasification.

Various types of reactors have been used to gasify biomass; they include fixed beds, entrained beds, moving beds, rotary kilns, and fluidized beds. Fixed bed and moving bed gasifiers have been used for quite a long time and a few commercial units are presently available. The other technologies are relatively new, and basically geared to large-scale operation; only very few have attempted commercialization. Reed and Jantzen (1980) summarized the state of gasifier research and the status of various gasifier manufacturers.

This introduction will only deal with the gasification of biomass in moving bed gasifiers of the downdraft type, with attention focused on research pertaining to some of the aspects of modeling and analysis of downdraft gasifiers. The modeling of another type of moving bed gasifier, namely the updraft type, has been reviewed in detail by Baekens and Schoeters (1985).

Moving bed gasifiers have been used to produce gas from biomass for more than a century. The first gas producer was built by Bischof in 1839; later in 1878, it was adapted for power generation purposes by Dowson (Wyer, 1906). Since then, numerous gas producers have been invented and the producer gas industry emerged. Producer gas continued to be the major gas supply until natural gas dominated the market in the 1930s; after that, almost all the producer gas plants were closed. However, the fuel shortages in World War II revived the interest in producer gas for a number of years. Due to the scarcity of liquid fuels in Europe during World War II, the search for domestically available fuels intensified and a great surge of activity in designing and using gas producers resulted. In Sweden, approximately 75,000 vehicles (40% of the automotive fleet) were converted to producer gas operation within two years. This experience was compiled by Swedish Academy of Engineering, and was translated by Reed and Jantzen (1979). After the war, the need for gasifiers dwindled and only a minimum level of research was maintained. The energy crisis in the past decade prompted a renewed search for alternative energy resources, and moving bed gasifiers once again obtained attention.

MOVING BED GASIFIER - DOWNDRAFT TYPE

Moving bed gasifiers generally fall into one of two types, updraft or downdraft, according to the flow patterns of solids and gas in the gasifier. In an updraft gasifier, the solid phase moves slowly downward, while the gaseous phase flows upward; the phases flow counter-current to each other. On the other hand, both the solid and gaseous phases flow downward in a downdraft gasifier; the phases flow co-current to each other. Downdraft gasifiers can be further classified into two groups according to their structure, one with a choke-plate and another without a choke-plate. Downdraft gasifiers with a choke-plate have air (or oxygen) injected in the choke region. Downdraft gasifiers without a choke-plate usually consist of a cylindrical column with the air (or oxygen) flowing through the top of the bed to the reaction zone; this type of gasifiers has been termed the 'stratified downdraft gasifier' by Reed and Markson (1982). Figures 1 and 2 (Reed, 1980) present schematic diagrams of typical updraft and downdraft choke-type gasifiers, respectively. Figure 3 (Reed and Markson, 1982) presents a schematic diagram of a stratified downdraft gasifier.

Although the mechanical features of a moving bed gasifier are simple, the physical and chemical processes in the reactor are extremely complex. For convenience, the active region of the reactor is ideally divided into a number of zones, as indicated in Figures 1-3. Due to the different flow patterns, the sequence of reaction zones in an updraft gasifier is different from that in a downdraft gasifier. With respect to the solids flow, the updraft type has the sequence pyrolysis-reduction-combustion, whereas the downdraft type has the sequence pyrolysis-combustion-reduction. Furthermore, the different flow patterns also result in significantly different performance

characteristics between the two types of gasifiers.

In an updraft gasifier, the solid phase moves in the direction of increasing temperature, while the gaseous phase flows in the direction of decreasing temperature. The solid phase is able to reduce to ash before leaving the system by reaction with incoming air (or oxygen). On the other hand, the volatile materials are carried by the gaseous phase into cooler regions and hardly have a chance for further gasification. Consequently, the product gas from updraft gasifier is rich in tar. Generally, this type of gasifier is only suitable for fuels with low volatile content, like coal or char.

In a downdraft gasifier, although the highest temperature is in the middle of the active region, the gas basically flows in the direction of increasing temperature; this enables the volatile materials from the pyrolysis zone to be cracked into permanent gases or combusted. However, the char can not react with oxygen; therefore, a certain amount of char output from the downdraft gasifier is unavoidable. The downdraft gasifier is usually suitable for fuels with high volatile contents, like biomass.

The main difference between the choke-type and stratified downdraft gasifiers is the location at which air (or oxygen) is injected. The choke-type introduces air in the middle of the reaction zone, whereas the stratified type introduces air at the top or both the top and the middle of the bed.

Reed and Markson (1982) measured temperature profiles in wood cylinders as they descended through a stratified downdraft gasifier. The temperature profiles are included in Figure 3. They concluded that the pyrolysis zone and combustion zone were indistinguishable, and therefore they combined them into a single zone, called the 'flaming pyrolysis' zone.

The choke-type downdraft gasifier has been used extensively for over a century; however, its scale-up has not been successful. The stratified downdraft gasifier was only developed recently and in an attempt to provide a basic design suitable for scale-up. The major limitation in the application of any moving bed gasifier is the size of the fuel. A typical dimension of 1-3 cm for fuels is required to provide adequate bed porosity to allow adequate flow of the gaseous phase and to avoid excessive pressure drop across the bed. Unlike fluidized bed gasifiers, downdraft gasifiers have limited means to control the ratio of feedstock to oxidizing agent. Instead, the inputs and outputs of the gasifier are managed to achieve steady-state operation, leaving the ratio of feedstock to oxidizing agent as an uncontrollable parameter.

EXPERIMENTAL STUDIES ON DOWNDRAFT GASIFIERS

A large amount of literature is available which describes various aspects of the design and operation of different types of downdraft gasifiers. However, very few studies have been published which contain complete mass and energy balances. Graham and Huffman (1981) investigated a commercial choke-gasifier rated at 1 MJ/hr. They reported complete mass balances, thermal efficiencies, mass conversion efficiencies and gas compositions with feed species, moisture content, and size and quality of feedstock as operating parameters. Walawender et al. (1985) reported on the performance of a commercial downdraft gasifier with a working capacity of 320 to 1,600 MJ/hr. Complete material balances as well as gas composition were reported over a wide range of throughput to examine the gasifier performance. These two studies represent the only works which present complete material balances for

moving bed downdraft gasifiers in the open literature.

MODELLING CONSIDERATIONS

Despite the use of downdraft gasifiers for biomass gasification for over a century, modeling attempts to describe this process have just started. Very few attempts have been made to model the entire gasifier. However, more fundamental studies, such as the analysis of the pyrolysis of small particles have been conducted to investigate the kinetics of biomass pyrolysis. The pyrolysis behavior of single large particles has also been investigated.

The reaction mechanisms involved in the pyrolysis of cellulosic materials are extremely complicated. It is believed that numerous elementary reactions take place in pyrolysis, in parallel and/or in series with each other. However, details on the kinetics of these reactions are lacking, because of the difficulty in isolating the individual reactions from each other. Shafizadeh (1968) proposed a conceptual mechanism for the pyrolysis of cellulosic materials which represented numerous experimental observations, as shown in Figure 4. As indicated in the figure, reactions 1, 2 and 3 refer to the primary reactions and reactions 4, 5, 6, 7 and 8 refer to the secondary reactions. The major product from the primary reactions is levoglucosan. The secondary reactions further reduce the products from the primary reactions (rich in tar) into lighter products (rich in permanent gases). When oxygen is present, some of pyrolysis products burn to produce H_2O and CO_2 and release heat.

There is an alternate conceptual chemistry often used to describe the gasification of biomass materials. The concept presents a qualitative

chemistry for pyrolysis by lumping the pyrolysis products into three major classes, char(solid phase), tar(liquid phase) and gas(gaseous phase). The virgin biomass is first thermally devolatilized to produce gas, char and tar.



Some of the tar is then further cracked in secondary reactions to become gas.



When oxygen is present, part of the product gas will burn to produce CO_2 and H_2O .



The combustion releases heat to maintain the temperature necessary for the devolatilization of biomass and secondary reactions. The char (consisting primarily of carbon) can then undergo gasification reactions with some of the constituents of the gaseous phase, such as H_2 , H_2O and CO_2 .



There are also reactions in the gaseous phase, such as the water gas shift reaction.



Two distinct solid-phase reactions are present in the downdraft gasifier, the pyrolysis of virgin biomass and the gasification of char. Virgin biomass is devolatilized in the pyrolysis zone to produce char and volatiles. The char is then further reduced in the reduction zone to produce additional gas. Due to the high volatile matter content of biomass (approximately 80%), the pyrolysis step is a major consideration. In this regard, the modeling of the pyrolysis of a single large particle is requisite to the modeling of a downdraft gasifier. The fundamentals of char gasification have been reviewed in detail by Graboski (1980).

Conceptually, the pyrolysis of a large biomass particle can be described by integrating (over the whole particle) the local intrinsic decomposition and the secondary reactions between the pyrolysis products. However, due to a lack of complete knowledge of the secondary reactions (in particular, tar cracking), it is natural that most of the models for the pyrolysis of large particles are based only on the kinetics of the decomposition step. The kinetics of the thermal decomposition of biomass have been explored primarily with small particles. A number of works have been conducted with small particles and these were recently reviewed by Milne (1980). It has been postulated that the decomposition of biomass can be represented by an Arrhenius-type kinetic expression, especially for cellulose. Since experimental data can usually be fit quite well with a range of adjustable parameters, first-order kinetics are often used for convenience. The activation energies determined for cellulose with this approach ranged from 26 to 60 kcal/gmole (Milne, 1980).

Due to this wide range of activation energies, it is suspected that a single-reaction kinetic model might not be adequate for the pyrolysis of cellulosic materials under all circumstances. Pitt (1962) suggested a multiple-reaction model with a distribution of activation energies for the pyrolysis of coal. This model assumed that there were many first-order parallel reactions competing with each other, and that the number of reactions was large enough to allow the use of a continuous probability function as follows

$$\int_0^{\infty} f(E) dE = 1 \quad (8)$$

where E designates the activation energy and $f(E)dE$ is the fraction of the reactions which have an activation energy from E to $E+dE$. The activation energy distribution functions, $f(E)$, is determined experimentally. Pitt used

this model to describe his thermogravimetric data for the devolatilization of coal. Furthermore, Raman et al. (1981) applied this model to describe the devolatilization of feedlot manure.

The heat of pyrolysis, while still controversial, is generally suggested to be endothermic, when the secondary gas phase reactions and char gasification reactions are not important.

Pyrolysis of a Single Large Particle

When a biomass particle of significant size is pyrolyzed, heat and mass transfer effects must be taken into account. Biomass is usually a poor heat conductor and, when subjected to heat, significant temperature variations may exist within the particle. On the other hand, the pyrolysis rate, at 250°C or higher, is considerably faster than the heat transfer rate. Whenever the temperature inside the particle reaches 250°C or more, pyrolysis initiates and begins to convert the solid phase into volatiles. The volatiles have a specific volume which is 2 to 3 orders of magnitude greater than the specific volume of the solid. Therefore, the large volume of the generated gaseous phase results in a vigorous gas stream bursting out of the particle at a significant velocity. The magnitude of this phenomenon can be illustrated from Reed and Markson's (1982) estimation that the velocity of gas emerging from a pyrolyzing particle subject to a relatively high temperature (900°C or higher) may be as high as several cm/sec. The escaping gases not only carry away a significant amount of sensible heat but they also hinder heat transfer to the interior of the particle, which limits the supply of the energy needed for the endothermic pyrolysis reaction. Consequently, the heat and mass transfer effects result in a wave like temperature profile traversing into the

particle. The pyrolyzing front driven by the impeded heat transfer moves inward with a growing char layer left behind. A schematic of the temperature and density profiles in a biomass particle during pyrolysis is shown in Figure 5. This conceptual model is supported by temperature and density profile measurements in a single biomass particle, e.g. Kannry, 1966; Maa and Bailie, 1973; and Belleville et al., 1982. The model leads to the conclusion that the pyrolysis of a large biomass particle is heat-transfer limited.

While emerging from the particle, the volatile materials generated by primary pyrolysis have a higher probability to further react within the gaseous phase or with the hot char layer in a large particle than they do in small particles. This phenomenon accounts for the differences in the end products of pyrolysis between large and small particles. For instance, Chan and Krieger (1983) reported the product distributions from the pyrolysis of a wood cylinder (radius 0.5cm) under fire-level radiation to be char 25.1%, tar 62.4% and gas 12.5%, whereas the product distribution from the pyrolysis of fine particles of white fir under fast heating (about 1000°C/s), reported by Brink and Mossondi (1978), was char 2.5%, tar 7.1% and gas 90%. Roy et al. (1983) studied the effect of particle size on the product distribution for the pyrolysis of wood. They reported the product distribution for the pyrolysis of aspen wood under slow heating conditions (10°C/min) for three different particle sizes, chips (about 2x2x0.5 cm), pins (about 1x0.3x0.3 cm) and wood flour. The product distribution for chips was oils 50.7%, water 16.3%, char 20.4% and gas 12.6%; for pins, oils 56.9%, water 13.9%, char 18.1% and gas 11.1%; and for wood flour, oils 60.8%, water 12.8%, char 15.8% and gas 10.6%.

Despite the complicated reaction mechanism, and the mass and heat transfer processes involved in the pyrolysis of a single large biomass particle, a number of models have been proposed to predict the pyrolysis behavior. These

models generally fall into two groups, one based the 'volume reaction' concept and the other based on the 'surface reaction' concept. Several representative models in each group are summarized in Table 1 and outlined as follows.

Volume Reaction Models

Bamford et al. (1946) proposed the first model for the pyrolysis of wood. This model considered the heat conduction of the solid, the heat of reaction and Arrhenius kinetics. Convective and radiative heat transfer were considered in the boundary conditions. The model satisfactorily predicted the central temperature versus time curve for the combustion of a wood slab.

Matsumota et al. (1968) extended Bamford's model to incorporate convective heat transfer within the particle with a temperature dependent thermal conductivity and applied it to the thermal decomposition of plastics. However, in their formulation they divided the space-time region into three distinct zones, namely the char, pyrolysis, and virgin material zone. Consequently, three sets of partial differential equations were used to describe the three zones. The oxidation of char at the particle surface was also included in the model which resulted in a recession of the char surface (particle surface). The prediction of the penetration velocity of the pyrolysis front (interface between the char and pyrolysis zones) and the recession velocity of char surface using this model was in good agreement with the experimental data. It was concluded that the temperature at the pyrolysis front was almost independent of the heat input at the particle surface and that it remained almost constant at 920°K over the course of the decomposition.

Kung (1972) developed a model for the pyrolysis of a wood slab, using density dependent thermal properties (specific heat and thermal conductivity).

The effects of slab thickness, char thermal conductivity, and the decomposition endothermicity on the pyrolysis rate were examined. It was found that (1) the pyrolysis rate of a thermally thick slab (in their case, $>1\text{cm}$) was quite sensitive to the char conductivity and endothermicity, while that of a thermally thin slab (in their case, $<0.02\text{cm}$) slab was insensitive to these parameters, and (2) a precise estimation of the thermal conductivity of char for the pyrolysis of the thermally thick slab was important due to its pronounced influence on the pyrolysis rate. Kung did not compare the results from his model with experiment due to the lack of suitable experimental data.

Havens et al. (1972) proposed a model for the pyrolysis of a large particle, which allowed the direct use of the data from differential scanning calorimetric and thermogravimetric measurements on small particles of the same material. As a result, no explicit kinetics were required and the heat of reaction was implied in the enthalpies of the solid and gaseous phases. The model was successfully applied to the prediction of the transient temperature distribution and volatile product evolution rates in the decomposition of wood cylinders. Application of their model required the following information (1) energy capacity data (total heat input required for a unit temperature increase per unit mass) obtained directly from a DSC, (2) weight loss data obtained directly from a TGA, and (3) experimentally measured values of the thermal conductivities for virgin wood and char at temperatures above 400°C . The requirement for extensive experimental data was the major disadvantage of their model and makes it semi-empirical.

Fan et al. (1977) proposed a mathematical model for pyrolysis, in which an effective diffusivity was used to account for the overall effects of molecular diffusion, eddy diffusion and convective flow. However, the authors pointed out that since pressure effects and size variations were neglected, the model

only applied to systems with a small fraction of pyrolyzable material under a slow heating conditions.

Kansa et al. (1977) incorporated the momentum balance into the modeling of wood pyrolysis. Their model included convective heat transfer between the solid and the gaseous phase, in contrast to the local thermal equilibrium assumed by other investigators. However, this effect was dropped out later in order to make numerical solution of the equations feasible. The predicted temperature distribution within a 5 cm slab almost coincided with the experimental data. The model also predicted the shape and magnitude of the pressure history within the particle, however the predicted pressure peak had a time lead of about 2 minutes compared to the experimental data. It was found that the pressure development within the particle was significantly influenced by the permeabilities of wood and char. The smaller the permeabilities, the greater the pressure build-up within the particle. When the permeability of char was small enough, oscillations (similar to a second-order damped response) in the pressure history as well as the pyrolysis rate were predicted by the model. This oscillatory phenomenon has not been observed experimentally.

A simple model for the pyrolysis of a wood cylinder was developed by Belleville et al. (1982). Surprisingly, only Fourier's law of heat conduction appeared in the energy balance with the thermal conductivity as a parameter. They experimentally measured the temperature in the center of cylinders as a function of time. Three different values of thermal conductivity were used over three consecutive temperature ranges. Their thermal conductivity values were determined by fitting the temperature-time curves. Consequently, the overall mass and heat transfer effects were lumped into the effective thermal conductivity parameter. The energy balance yielded the time-resolution

temperature distribution inside the particle. With the temperature distribution determined, the density distribution was then calculated using first-order kinetics. This model predicted the weight loss versus reaction time quite well.

The pyrolysis of cellulosic materials usually proceeds with some size reduction. This effect, ignored by most investigators, was incorporated in a recent model proposed by Villermaux et al. (1983). The particle was assumed to have a uniform density profile over the course of pyrolysis with the local ratio of volatiles to char varying. As the volatiles escaped, the particle was required to shrink to maintain a uniform density profile. Consequently, the remaining char had a density equal to the virgin solid. This behavior contradicts the experimental observation that the density of char is considerably lower than that of virgin solid. Consequently, this model can not be applied to materials with a high volatile matter content, like biomass.

All of the models discussed so far have focused on the time for complete pyrolysis or the influence of heat transfer rate or the rate of weight loss. No effort has been made to incorporate the secondary reactions or to predict the product distribution. It is generally recognized that the relative amounts of volatiles and char vary with the heating history, the nature of the secondary reactions and the particle size (see, e.g., Roberts, 1970). Chan and Krieger (1983) recently proposed a model for the prediction the relative amounts of gas, tar and char. The primary decomposition reaction as well as secondary reactions of the gases and tars were taken into account in their model. The model predicted the temperature profile within the particle and the relative amounts of char, tar, and gas for the pyrolysis of wood slab fairly well.

Surface Reaction Models

The unreacted shrinking core model was first applied to the pyrolysis of cellulosic materials by Maa and Bailie (1973). Decomposition was assumed to proceed only at the boundary between the char layer and the unreacted core, while the temperature within the core was assumed to be uniform and the same as that of reacting boundary. The boundary moved slowly towards the center of the particle as pyrolysis progressed. The pseudo-steady state assumption was imposed to facilitate numerical solution. One advantage of this model was the ease of solution of the system of equations, which were reduced to ordinary differential equations by the use of pseudo-steady state assumption. A sensitivity analysis was conducted to determine the effects of the model parameters. The most significant parameters were found to be the wall temperature, activation energy, frequency factor, effective thermal conductivity, heat of reaction and solid density. They indicated that pyrolysis was controlled by two competing mechanisms, the heat transfer rate and kinetic reaction rate. They proposed a criterion to determine the controlling mechanism based on the particle size and wall temperature. For a large particle (in their case, > 3 cm), pyrolysis was in the heat transfer controlled region, and the time for complete pyrolysis (τ) was estimated (in their case) from

$$\tau = r^2 \rho \Delta H / (240 k_e (T_e - T_c)) \quad (9)$$

where r , ρ , ΔH , k_e , T_e and T_c represent the radius of the particle, the density of solid, the heat of reaction, the effective thermal conductivity, the surface temperature of the particle and the temperature of the unreacted core. For a small particle (in their case, < 0.1 cm), pyrolysis was in the reaction controlled region and the time for complete reaction was estimated (in their case) from

$$\tau = r / (60k_0 e^{-E/RT}) \quad (10)$$

where k_0 , E , R , and T represent the frequency factor, activation energy, gas constant and local temperature. For a particle of intermediate size, both heat transfer and reaction rates need to be considered.

An interesting 'Phase-change' model, aimed at simplifying the mathematics of modeling the pyrolysis of cellulosic materials, was proposed by Derosiers and Lin (1983). The main feature of their model was that a kinetic expression was not needed. The solid was assumed to decompose into char and volatiles immediately when its temperature reached a prescribed temperature, which was treated as an experimentally determined parameter. As the particle was heated, a phase change like boundary shifted toward the center of the particle. Behind the boundary only a char layer remained, while ahead of the boundary temperature variation within the virgin solid was allowed. The model predicted the central temperature and weight loss of the particle versus time.

Unavoidably, most these models include complicated mathematical equations; coupled mass and energy balance partial differential equations and/or kinetic expressions. The effort required in obtaining numerical solutions is often quite formidable. Still, considerable experimental work is needed in order to independently determine the parameters required for a specific model. These models may find very limited application for design and prediction purposes, especially in real gasifiers. For example, in a moving bed gasifier, the properties of the gaseous phase (i.e. the environment surrounding the particles) vary significantly within the gasifier. These properties include the temperature, composition, velocity and pressure (although pressure effects are often negligible). Consequently, the modeling of a moving bed gasifier is much more complicated than for a single particle.

Modeling of the Downdraft Gasifier

Although the downdraft gasifier has been used for the gasification of biomass for over a century, modeling efforts are just beginning. The behavior of the particle and gas phases combine to give the performance of a gasifier. However no modeling work for the entire moving bed gasifier, based on the single particle model, has been successful. To date there has been only one single particle model (Derossiers and Lin; 1983) ever used to estimate the temperature profile in the pyrolysis zone of an updraft gasifier.

A recent review by Bukenens and Schoeters (1985) summarized the modeling efforts for moving bed gasifiers. However, the models were all based on updraft gasifiers with coal and low volatile content materials as the fuel. The behavior of a downdraft gasifier is quite different from that of an updraft type. The major reaction for biomass fuels is pyrolysis, whereas for coal and other low volatile content materials it is gasification. These basic differences make the models for updraft gasifiers not readily applicable to the downdraft type.

The work of Reed and Markson (1982) seems to be the first work to attempt to directly model a downdraft biomass gasifier. Their model divided the active portion of the gasifier into two major sections, a flaming pyrolysis zone and a char reduction zone. They based their analysis of the flaming pyrolysis zone on Huff's (1985) empirical equation, which correlated the time required to complete flaming combustion of a particle with temperature, particle size, shape, density and moisture content. They modified Huff's equation to incorporate the depletion of oxygen and developed an equation for estimating the time to complete flaming pyrolysis at a given temperature. In a gasifier, the temperature as well as oxygen concentration changes with axial

position; therefore, an experimentally determined correction factor was required to determine the flaming pyrolysis time. With a given solid flow rate, the length of the flaming pyrolysis zone could then be determined. For the char reduction zone, the time and hence bed length was determined using first-order kinetics for the Boudouard reaction.

Reed et al. (1983) proposed a mathematical model to predict the temperature and composition of the product exiting pyrolysis zone, and to simulate the temperature and composition profiles of the gas and solid phases in the char reduction zone. The output from the pyrolysis zone served as the input for the char reduction zone.

CONCLUDING REMARKS

This brief review is aimed at improving the understanding a particular biomass gasification processes; downdraft gasification. Special attention has been given to some of the modeling aspects. Due to the transport phenomena limitations, only biomass of the macro-particle size is suitable for gasification in the downdraft gasifier. As a precursor to gasifier modeling, models for a single large particle are required. However, these models are often too complex to be used in the mathematical modeling of the complete gasifier. In this regard, experiments are always required to correlate the model predictions with the operating parameters of a specific gasifier. Well developed models for prediction and design purposes are still lacking. Actually, the gasifiers that are in operation today have been designed by experience.

A large number of studies have concentrated on various aspects of the

design and operation of different downdraft gasifiers. Very few experimental results have been reported concerning the modeling aspects of a general gasifier. Those experimental results reported have focused only on the mass and energy balances. To date, no work has been attempted to investigate the influence of momentum transfer in a downdraft gasifier in which the flowrates of gases and solid, temperature and the size of solids change with position.

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Table 1. Summary of Single Large Particle Pyrolysis Models

Author	Material Pyrolyzed	Energy Balance	Type of Rxn. model	Kinetica of Pyrolysis	Comment
Bamford et al.(1946)	wood slab	a,b,c	volumetric	first order	
Matsumoto et al.(1968)	plastic slab	a,b,c,d	'	'	3-zone
Kung(1972)	wood slab	a,b,c,d	'	'	
Havens(1972)	wood cylinder	a,b,c,d	'	no explicit kinetics	
Fan et al.(1977)	coal(general shape)	a,b,c,d	'	n-th order	
Kansa et al.(1977)	wood slab	a,b,c,d	'	first order	momentum balance
Belleville et al.(1982)	wood cylinder	a,b	'	'	
Villiermaux et al.(1983)	biomass slab	a,b,c,d	'	'	particle shrinking
Chan and Krieger(1983)	Biomass pellet	a,b,c,d	'	first order (multi-rxn)	
Mao and Bailie(1973)	cylinder	a,b,c,d	surface	first order	2-zone
Derosiers and Lin(1983)	wood cylinder	a,b,d	'	none	2-zone

a: transient response; b: conduction; c: heat of reaction; d: convection.

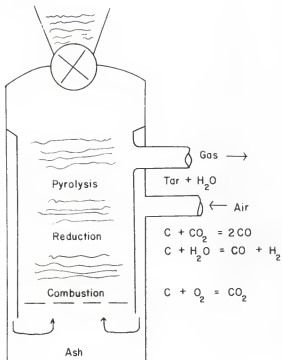


Figure 1. Schematic Diagram of Updraft Gasifier (Reed, 1980).

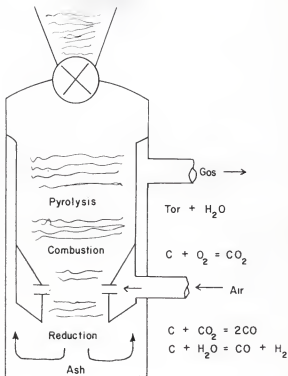


Figure 2. Schematic Diagram of Downdraft Choke-Gasifier (Reed, 1980).

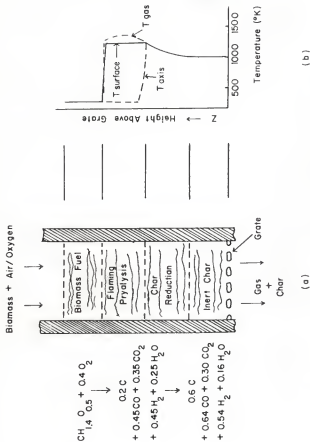


Figure 3. Schematic Diagram of a Stratified Downdraft Gasifier with Representative Temperature profiles (Reed and Marksan, 1982).

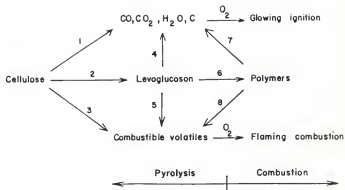


Fig 4. The pyrolysis and combustion of cellulose.

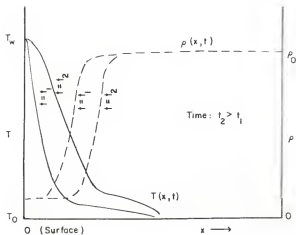


Figure 5. Schematic Diagram of Time-Resolution Temperature and Density Profiles for the Pyrolysis of a Large Biomass Particle.

CHAPTER 3

WOOD CHIP GASIFICATION
IN A COMMERCIAL DOWNDRAFT GASIFIER

INTRODUCTION

Fixed-bed and moving-bed gasifiers for the production of low Btu gas from wood and charcoal were widely used in Sweden and other countries during the World War II era. The Swedish experience was compiled by the Swedish Academy of Engineering, and this work was recently translated by Reed and Jantzen (1979). After the war, the need for gasifiers dwindled although the Swedes continued their development efforts.

Following the Arab oil embargo, rapidly rising prices for petroleum coupled with projections of potential near-term petroleum and gas shortages prompted the search for alternative energy sources. In response to this effort, gasification technology has been resurrected and low BTU gas production became one of the major options for alternative energy in the United States. Reed et al. (1980) reviewed the state of the art in gasification technology and summarized the status of a number of commercial and research efforts in the field.

Although the literature contains a vast amount of information on various facets of the gasification of wood and biomass in fixed-bed and moving-bed gasifiers, very little has been published on the complete material balance for these gasifiers. In fact, many gasifiers have been constructed with little or no quantitative information on the gasifier performance. Graham and Huffman (1981) presented detailed information on the Biomass Corporation gasifier; this appears to be the only complete material balance analysis of a moving bed gasifier in the open literature.

The present work reports the material balances and performance of a recently commercialized downdraft gasifier, the Buck Rogers 'Gasafire'™. The

gasifier, rated to produce 320 to 600 MJ/hr, was evaluated over its working capacity range and was found to be remarkably consistent in terms of gas yield, gas composition, gas heating value and other measures characterizing its performance.

EXPERIMENTAL FACILITIES AND PROCEDURE

Gasifier Description

Figure 1 presents a schematic representation of the commercial gasifier used in this study. It is unique in that there is no throat or choke plate to support the bed. A fan located downstream from the gasifier draws air through the chip bed as well as the tuyeres. A feed bin and a screw feeder (not shown in Figure 1) supply chips to the top of the gasifier each time the bed height drops to a set level, controlled by an electric eye. The fan motor also drives the cyclone rotary valve and the 'airgitator' gearbox through a series of speed reducing pulleys and gears. Consequently, the fan speed controls the air intake, chip feed, gas production and char output rates. The 'airgitator' rotates very slowly at 20 to 30 minutes per revolution and has several functions, listed below:

- (1) It provides secondary air, through the tuyeres.
- (2) It levels the feed at the top of the bed.
- (3) It slowly mixes material throughout the gasifier.
- (4) It generates a grinding motion in the ceramic ball bed which, in turn, reduces the char size so that it can percolate through the ball bed and finally through the grid plate.

(5) Coupled with ball bed and the grid plate, it serves as the support for the bed.

Gas and char flow out of the gasifier through the grid plate. The fan suction draws them out of the base of the gasifier and blows them through the cyclone where the char is separated from the gas. The gas stream is then sent to a flare stack (not shown in Figure 1) where it is incinerated.

Operating Procedure

The procedure used in operating the gasifier is outlined below.

1. Start-up. The fan was started at low speed, and propane was introduced into the flare stack and ignited. Next the feeder was manually operated to fill the gasifier just above the tuyeres. The bed was then ignited with an electric fire starter and the fan speed was increased. The fire was allowed to develop for 15 to 20 minutes. Then chips were added to fill the gasifier and the fan speed was increased. As gas production increased, the propane rate was decreased and finally shut down. The fan speed was then gradually increased to the desired operating condition and the reactor was allowed to stabilize. This procedure normally required two hours to complete.
2. Char and feed measurements. Char was measured by direct collection over 30 minute intervals. Chip feed was measured by a procedure to be detailed later. Both of these measurements were conducted over a five hour period starting after the completion of step 1.
3. Gas analysis and condensables measurement. One hour after the char and feed measurements were started, a small gas stream was drawn for gas analysis and condensate determination. The specific procedures will be detailed later.

These measurements were conducted for two hours.

4. Tracer gas measurement. Nitrogen tracer gas was introduced to the product gas at a known rate, and the gas analysis and condensate measurements were repeated. The specific procedures are detailed later. These measurements were conducted for one hour.

5. Post tracer measurements. Step 3 was again repeated without the tracer gas for a one hour period. 6. Shut-down. The chip feed was stopped and the fan speed was increased to accelerate complete consumption of the chips. This normally required one hour.

Supporting Measurements

The supporting measurements included moisture and ash analyses of the feed and elemental analysis of the feed. Moisture was determined in a muffle furnace by standard ASTM procedure. The elemental analysis of feed was conducted with the aid of a Perkin-Elmer Model 240b Elemental Analyzer.

Feedstock

The feed material used in this work was a mixed hardwood chip. The average chip screen size was 14 mm; the approximate size distribution is given in Table 1. The chips were free of bark and were pre-dried to 6 - 17% moisture (wet basis). The elemental analysis of the dry chips was fairly constant over the course of the experiments. The average analysis was C, 48.31%; H, 5.94%; O, 44.77%; N, 0.18%; ash, 0.80%. The average higher heating value of the dry chips, estimated from the Dulong formula, was 16.82 MJ/kg.

Measurement Scheme and Computations

The objective of the present work was to gather complete material balance data for the gasifier and to evaluate the gasifier performance. Because of the configuration of the gasifier and the stream flow rates involved, it was not convenient to measure all of the stream rates directly; consequently, indirect methods were devised.

The two quantities directly measured were the char output rate and the dry gas composition. Char output was determined by collecting and weighing the char over 30 minute intervals for approximately a five hour period. The gas analysis was conducted with an Applied Automation on-line process gas chromatograph (GC) which drew a continuous sample from points downstream of the cyclone. The GC had a cycle time of about 11 minutes and was able to detect the following components in the producer gas: H_2 , CO_2 , CO , CH_4 , O_2 and N_2 along with traces of C_2H_6 , C_2H_4 and C_3H_4 . It was believed that the O_2 in the producer gas resulted from air leaks in the fan seal and the gas sampling system. Consequently all compositions were converted into an air-free basis. Product gas compositions were determined from the average of the pre and post trace gas analyses.

The dry product gas rate was indirectly determined with the aid of a tracer technique. Nitrogen tracer was introduced into the product gas at a known volumetric rate at a point just upstream from the fan. The gas composition was then measured with the tracer flowing over a one hour period to insure steady state compositions. From the known tracer, the dry gas volumetric flow rate could be readily evaluated. Knowing the molecular weight of the gas allowed the stream rate to be converted to a mass basis.

The dry gas flow rate, coupled with a nitrogen balance between the inlet and outlet of the gasifier, allowed the dry air input rate to be evaluated. This balance assumed that negligible nitrogen was produced in the gasification process. Knowing the temperature and relative humidity over the course of an experiment allowed the moisture input from humid air to be evaluated. Since most experiments were of five hour duration, the moisture evaluation was based on mid-run conditions.

Condensables consisted of two parts, tar and aqueous. These were determined by taking a side draw from the main gas stream at a rate of about $0.55 \text{ m}^3/\text{hr}$. This stream also supplied the GC with sample. The gas was first sent through a hot filter (maintained at about 420°K with a heating tape) which was packed with glass wool to remove heavy tar components. Next the gas was sent through two water cooled condensers in series to remove most of the water. The remaining water removal was accomplished with two receivers placed in series in an ice bath. The tarry mist was then removed by passing the stream through a tightly packed glass wool filter. Flow through the train was maintained with the aid of a 'Gast' compressor which provided suction. The discharge from the compressor was then sent through a wet test meter, followed by a drierite chamber and then to the GC. Two identical trains were constructed for use with and without tracer. The wet test meter readings were corrected for temperature, pressure, water of saturation and air, and tracer gas when needed. The condensates and tars were weighed and converted to a mass per unit volume dry gas basis. These results were then scaled to the main stream flow.

The chip feed to the gasifier was intermittent rather than continuous. However, chip flow through the gasifier was continuous with only the chip level fluctuation ($0.3 - 0.5 \text{ m}$). Chips were delivered by a constant speed

screw feeder which was activated by the chip level sensed by an electric eye. The following procedure was adopted to evaluate the chip feed rate. A timer was wired into the screw feeder. This was generally recorded over a five hour period. The feeder was calibrated by collecting and weighing the feeder discharge over 5 to 10 sec intervals. Some fluctuations in the calibration were noticed, i.e., 0.6 to 0.7 kg/sec. These were due to bin height variations and/or bulk density variations. The feed rate for each run was evaluated by taking the appropriate calibration factor times the cumulative on-time for that run. The above items determined the material balance for the gasifier. Material balance closure was evaluated as output/input and the results adjusted so that all closures were less than 100%, i.e., for closure greater than 100% the result was subtracted from 200.

The performance of the gasifier was evaluated in terms of a variety of measures determined from the gas composition, material balance data and supporting measurements. The dry gas composition gave one measure of gasifier performance and from it, the higher heating value of the dry gas was calculated from the standard heat of combustion of the components. The gas, char and tar yields gave other measures of performance. These were calculated on the basis of a unit mass of dry feed to eliminate the influence of moisture variations. A dry ash-free basis was not used since the ash content of the dry chips was less than 1% and not subject to much variation. Char yield was expressed as the kg of char per 100 kg of dry chips and gas yield was expressed on a volumetric basis as m^3/kg of dry chips. The volume basis used in this work was 289°K and 101.3 kpa. Tar yield was expressed in two ways, one as kg of tar per 1000 kg of dry chips and the other as parts per million in the dry gas. Also evaluated were the water output, kg of condensate per kg of dry chips. The energy output was evaluated in MJ/hr and the energy yield in MJ/kg of dry chips. The cold gas efficiency was evaluated as the ratio of

the energy content of the gas produced from one kg of dry chips to the energy content of one kg of dry chips. This quantity represented the fraction of the energy content of the feed that was converted to combustible gas. The mass conversion efficiency was evaluated as the ratio of dry gas output to the total input.

RESULTS AND DISCUSSION

Results

The material balances for runs with closures greater than 95% are summarized in Table 2. A total of 19 runs were completed, and 9 had closures greater than 95%; the poorest closure was 80%. Some runs had closures greater than 100% and these were adjusted to a 100% basis by subtracting the result from 200. The material balance closures for all runs are included in Table 3.

Table 3 presents various measures of the gasifier performance as a function of the dry chip feed rate along with the moisture content of the as-received feed. Included in the table are the char yield in kg/100 kg of dry chips, the tar yield in kg/1000 kg of dry chips, the gas higher heating value, the energy yield per unit mass of dry chips, the total energy output, the water output rate in kg/kg of dry chips, the tar content of the gas in ppm, the dry air input rate in kg/kg of dry chips, the volumetric gas yield, the mass conversion efficiency and the cold gas efficiency. The means and standard deviations for the complete data set and the nine runs with closures greater

than 95% are given at the bottom of the table. As can be seen from the table, although the values fluctuated somewhat, no distinct trends were observed as the dry chip feed rate increased. The mean gas yield over the entire operating range was 2.19 m³/kg for runs with closures greater than 95%. Other important performance measures include the mean heating value at 5.51 MJ/m³, the mean mass conversion efficiency at 88%, and the mean cold gas efficiency at 72%.

The energy yield per kg of dry chips was fairly close for the runs with closures greater than 95%. This behavior is also illustrated by Figure 2 in which the total energy output is plotted against the dry chip feed rate. As can be seen from the figure, a linear relationship determined by least squares analysis describes the data quite well.

Dry gas composition data as a function of dry chip feed rate for all the runs are summarized in Table 4. The principal components of the product gas were H₂, CO, CO₂, CH₄ and N₂ with trace amounts of C₂H₄, C₂H₆ and C₃H₈. Although the results show some fluctuation, there is no distinct trend with increasing dry chip feed rate. As noted earlier, all gas compositions were determined from the average of the pre and post trace gas analyses and are reported on an air-free basis.

Discussion

The results of this work indicate that the performance of the gasifier is remarkably consistent over a fairly wide range of operating conditions. The chip feed rate was varied over almost a five fold range from 27 to 126 kg/hr with a corresponding energy output ranging from 320 to 1400 MJ/hr. The

consistency in performance was further reflected by the temperature measured right above the ceramic ball bed. It fluctuated only slightly from 950 to 1000°K over the course of the experiments with no distinct trends indicated as the chip feed rate was increased.

The mass conversion and cold gas efficiencies obtained in this study are compatible with the results obtained by Graham and Huffman (1981) using a downdraft gasifier of comparable capacity. Using poplar chips with 13% moisture content (wet basis), they reported a mass conversion efficiency of 88% and a cold gas efficiency of 74%. Their feed rates were comparable to the upper range examined in this work, but they made no effort to study wide ranges of feed rate. Their gas compositions were similar to ours but showed more CO (24 - 26%) and less CO₂ (9 - 10%). The differences are probably due to the differences in composition of the feedstocks used in the two studies. Furthermore, Graham and Huffman (1981) reported substantially more tar than we did, i.e., of the order of 10,000 ppm. Their tar results were determined by difference whereas ours were measured values. Our measurements of tar were rather crude because some of the tar was deposited on the sample system surfaces and not measured. Even if the measurements were low by 100%, the tar observed in our study was still an order of magnitude lower than that reported by Graham and Huffman.

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Table 1. Chip Size Distribution

Screen size passed (mm)	Weight fraction	Typical sizes (mm)
sticks	0.014	10x75, 7x74, 23x95
> 18.5	0.099	24x43, 21x32, 30x47
18.5	0.211	17x47, 20x25, 15x26
12.5	0.592	15x32, 12x17, 9x31
4.5	0.056	5x19, 3x41, 4x7
3.5	0.028	1.5x13, 2x9, 2x2

Table 2. Material Balance Summary

Run	Inputs (kg/hr)			Outputs (kg/hr)					Overall	
	Wet chips	Dry air	H ₂ O	Total	Dry gas	Char	Tar	H ₂ O	Total	Closure(%)
98	32.0	43.1	0.5	75.6	66.3	0.9	0.09	7.4	74.7	98.9
910	32.0	45.2	0.5	77.7	68.0	0.9	0.14	7.4	76.4	98.3
920	35.7	62.1	0.4	98.2	92.9	1.4	0.09	7.1	101.5	96.6
101	52.4	76.9	0.8	130.1	116.9	1.7	0.09	13.3	132.0	98.6
929	52.7	74.0	1.0	127.7	113.0	1.8	0.14	12.2	127.1	99.5
106	58.1	76.8	0.5	135.4	117.0	1.2	0.09	10.8	129.1	95.4
1119	89.1	112.1	1.0	202.2	173.8	3.0	0.27	19.7	196.8	97.3
1117	96.2	140.4	0.7	237.3	218.9	2.5	0.18	22.9	244.5	97.0
1221	140.5	202.1	1.1	343.7	302.1	4.1	0.54	43.3	350.0	98.2

Table 3. Performance Summary

Run	Dry Chip Rate(kg/hr)	Chip Moisture (% wet basis)	Matl. Balance Closure(%)	Dry Gas HHV (MJ/m ³)	Gas Yield (m ³ /kg DC*)
98	26.7	16.52	98.9	5.33	2.22
910	26.7	16.52	98.3	5.18	2.28
920	33.6	6.01	96.6	5.33	2.48
927	41.2	8.11	85.4	5.63	2.54
922	47.8	5.75	85.1	5.44	1.43
101	48.7	6.99	98.6	5.59	2.18
929	48.8	7.41	99.5	5.63	2.10
1011	49.5	7.31	91.6	5.44	2.55
1015	52.3	7.95	88.8	5.33	2.73
924	53.7	5.50	91.8	5.66	1.62
106	54.3	6.67	95.4	5.48	1.95
123	54.3	14.61	94.2	5.51	1.90
104	63.7	7.72	92.6	5.55	2.61
1129	68.9	14.27	91.9	5.59	2.85
121	72.5	15.07	80.4	5.63	1.39
1119	74.6	16.32	97.3	5.74	2.11
1117	85.4	11.26	97.0	5.78	2.30
1110	119.7	11.56	91.0	5.74	1.79
1221	125.5	10.67	98.2	5.44	2.10
all data					
mean	—	—	—	5.51	2.17
σ	—	—	—	0.15	0.41
>95%					
mean	—	—	—	5.51	2.19
σ	—	—	—	0.19	0.14

*DC represents Dry Chips.

Table 3. Continued

Run	Char Yield (kg/100kg DC)	Tar Yield (kg/10 ³ kg DC)	Tar in Dry Gas (ppm)	Dry Air In (kg/kg DC)	H ₂ O Out (kg/kg DC)
98	3.28	2.55	1370	1.61	0.277
910	3.21	4.59	2003	1.69	0.277
920	4.06	2.71	977	1.85	0.212
927	3.08	3.30	1180	1.75	0.438
922	2.76	1.90	1199	1.05	0.186
101	3.45	1.86	777	1.56	0.273
929	3.72	2.79	1205	1.52	0.250
1011	3.94	4.59	1607	1.88	0.321
1015	4.69	5.21	1679	2.08	0.381
924	2.79	2.54	1426	1.13	0.209
106	2.18	1.67	776	1.42	0.199
123	4.27	3.35	1594	1.35	0.236
104	2.71	2.85	989	1.90	0.307
1129	3.82	3.96	1254	2.04	0.281
121	3.57	3.76	2445	0.99	0.169
1119	3.96	3.65	1567	1.50	0.264
1117	2.92	2.13	830	1.64	0.268
1110	1.67	3.03	1544	1.27	0.206
1221	3.25	4.34	1803	1.61	0.345
all data					
mean	3.33	3.20	1380	1.57	0.268
σ	0.72	0.99	424	0.31	0.067
>95%					
mean	3.34	2.92	1256	1.60	0.263
σ	0.54	0.99	433	0.12	0.039

Table 3. Continued

Run	Energy Yield (MJ/kg DC)	Total Energy Out (MJ/hr)	Mass Conversion Efficiency (%)	Cold Gas Efficiency (%)
98	11.82	315	87.7	70.2
910	11.82	315	87.5	70.2
920	13.14	441	94.7	78.1
927	14.31	590	98.1	85.0
922	7.76	370	74.9	46.1
101	12.19	594	89.8	72.5
929	11.82	576	88.5	70.2
1011	13.84	685	96.1	82.2
1015	14.52	759	97.5	86.3
924	9.11	490	80.9	54.2
106	10.68	579	86.5	63.5
123	10.44	567	83.0	62.1
104	14.50	924	96.2	86.2
1129	15.89	1094	98.1	94.4
121	7.81	567	70.8	46.5
1119	12.08	901	86.0	71.8
1117	13.31	1136	92.2	79.1
1110	10.23	1224	81.6	60.8
1221	11.42	1434	87.9	67.9
all data				
mean	11.93	---	88.3	70.9
σ	2.25	---	7.5	13.0
>95%				
mean	12.03	---	87.9	71.5
σ	0.76	---	2.1	4.5

Table 4. Dry Gas Composition (%)

Run	H ₂	C ₂ H ₄	CO ₂	C ₂ H ₆	C ₃ H ₈	N ₂	CH ₄	CO
98	15.00	0.10	16.67	0.39	0.07	46.92	2.50	18.41
910	15.00	0.08	16.25	0.36	0.06	47.89	2.33	18.05
920	14.27	0.09	15.50	0.42	0.07	48.19	2.49	18.97
927	16.56	0.08	16.60	0.41	0.07	44.45	2.52	19.31
922	14.85	0.09	15.24	0.42	0.07	47.46	2.48	19.40
101	15.47	0.10	14.80	0.49	0.09	46.72	2.41	19.94
929	15.52	0.09	14.96	0.47	0.07	46.50	2.55	19.85
1011	14.69	0.12	16.28	0.57	0.09	47.55	2.61	18.09
1015	13.25	0.14	16.31	0.65	0.10	49.14	2.67	17.73
924	16.04	0.08	15.84	0.44	0.07	45.11	2.51	19.93
106	15.32	0.07	15.23	0.39	0.07	46.78	2.45	19.69
123	16.33	0.11	16.81	0.57	0.09	45.94	2.39	17.77
104	15.25	0.08	14.83	0.45	0.06	46.83	2.48	20.01
1129	15.61	0.12	15.90	0.57	0.09	46.22	2.41	19.07
121	16.06	0.13	16.41	0.63	0.10	45.86	2.47	18.34
1119	15.56	0.14	15.51	0.63	0.09	45.96	2.49	19.62
1117	14.90	0.15	15.66	0.66	0.10	46.02	2.61	20.00
1110	15.50	0.13	15.17	0.58	0.09	45.95	2.49	20.08
1221	12.25	0.19	16.24	0.82	0.10	49.40	2.81	18.20

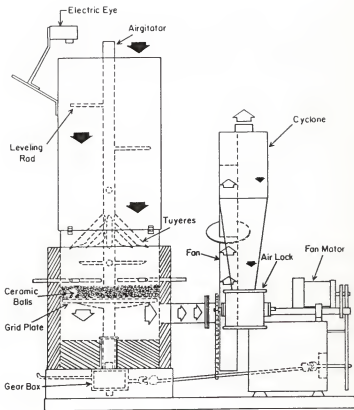


Figure 1. Commercial Gasifier

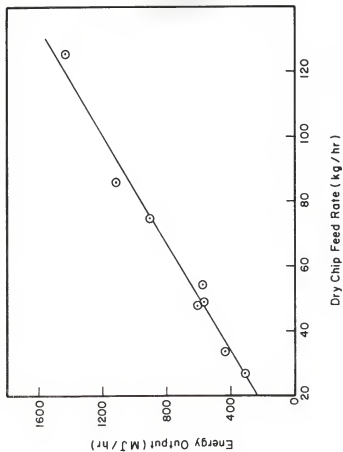


Figure 2. Energy Output Versus Dry Chip Feed Rate

CHAPTER 4

MASS AND ENERGY BALANCE ANALYSIS

ON A DOWNDRAFT GASIFIER

INTRODUCTION

Moving bed gasifiers have been used to produce gas from biomass for more than a century. The first gas producer was built by Bischof in 1839; later in 1878, it was adapted for power generation purposes by Dowson (Wyer, 1906). Since that time, numerous gas producers have been invented and for a time, the producer gas industry grew rapidly. Producer gas was the major gas supply until natural gas dominated the market in the 1930s; after that, almost all the producer gas plants were closed. However, the fuel shortages in World War II revived interest in producer gas for a number of years. Due to the scarcity of liquid fuels in Europe during World War II, the search for domestically available fuels intensified and a great surge of activity in gas producers resulted. In Sweden, approximately 75,000 vehicles (40% of the automotive fleet) were converted to producer gas operation within two years. Their experience was compiled by Swedish Academy of Engineering, and was translated to English by Reed and Jantzen (1979). After the war, the need for gasifiers dwindled and only a minimum level of research work was maintained. The energy crisis in the past decade prompted a renewed search for alternative energy resources, and moving bed gasifiers once again obtained attention.

A large amount of literature is available which describes various aspects of the design and operation of different types of downdraft gasifiers. However, very few studies have been published which contain complete mass and energy balances. In fact, many gasifiers have been constructed on a trial and error basis. Graham and Huffman (1981) investigated a commercial-scale choke-gasifier rated at 1 MJ/hr. They reported complete mass balances, thermal efficiencies, mass conversion efficiencies and gas compositions with the feed species, moisture content, and the size and quality of feedstock as

operating parameters. Walawender et al. (1985) reported on the performance of a commercial downdraft gasifier with a working capacity of 320 to 1,600 MJ/hr. Complete material balances as well as gas compositions were reported over a wide working range to examine the gasifier performance. These two studies represent the only complete material balance analyses of moving bed downdraft gasifiers in the open literature.

A few studies have been published concerning the energy analysis of pyrolysis processes. Otoma and Gotch (1979) applied the second law of thermodynamics to the analysis of a pyrolysis system. Shieh and Fan (1982) developed formulas for estimating the energy and available energy contents of structurally complicated materials, such as coal and biomass. Recently, rigorous mass, energy and available energy balances for a conceptual pyrolysis system were developed by Ishimi et al. (1982). However, at present, no detailed thermodynamic analysis has been reported on the moving bed downdraft gasifier.

This work presents the complete material and energy balances for a commercial moving bed downdraft gasifier. It is based on the experimental data of Walawender et al. (1985) for a commercial moving bed downdraft gasifier. One objective of this work was to formulate an empirical stoichiometry to represent the overall gasification process. A second objective was to conduct a thermodynamic analysis to determine the efficiency of the process; the approach was based on applications of the first and second laws of thermodynamics.

MASS BALANCE ANALYSIS

The objective of the material balance analysis was to formulate an empirical stoichiometric equation for the gasification of wood chips based on the data of Walawender et al. (1985). The formulation of the stoichiometry was based on an analysis of the experimental overall material balance data for the gasifier and the development of elemental material balances for the process through a two level adjustment of the data to force closure.

To formulate the stoichiometry for the overall gasification reaction requires a perfect balance for all of elements involved in the reaction. Therefore, if a set of experimental data with perfect balances for all elements exists, the empirical stoichiometry will result naturally. However, perfect closure is rare for a real process, especially one as complicated as a moving bed gasifier. The inherent variability of biomass, measurement errors and system fluctuations all contribute to make complete closure impossible.

The material balance closure, both overall and elemental, is used in this analysis as a measure of how well material balance has been achieved. It is defined as the mass ratio of output to input. Nine runs from the work of Walawender et al. (1985) were selected for the initial analysis. All had overall closures to within 5%. The data base is summarized in Table 1. However, one of the nine runs, Run 101, was rejected due to its poor elemental balance closures in the preliminary analysis.

Table 2 presents the overall material balance closures and the elemental balance closures for the remaining runs. Closures were evaluated based on the following assumptions.

(1) The elemental composition of dry chips was based on the mean of the

- available experimental data. This composition is presented in Table 3.
- (2) Dry air consists of oxygen and nitrogen only, with a molar ratio of 21 to 79. Its composition is included in Table 3 on a weight basis.
- (3) The elemental composition of tar was assumed to be the same as dry chips due to the lack of experimental data. Although this introduces significant error in the elemental composition of tar, its effect on the material balance is negligible due to the extremely small amount of tar compared to other streams (see Table 1).
- (4) The elemental composition of char was assumed to be the mean of the available experimental data. This composition is included in Table 3.

As can be seen from Table 2, the carbon(C) closures on the original data ranged from 0.82 to 1.09, hydrogen(H) closures ranged from 0.84 to 1.02, oxygen(O) closures ranged from 0.98 to 1.10, all nitrogen(N) closures were 0.99, and overall closures ranged from 0.95 to 1.03. Among the elemental balance closures, the poorest were those for hydrogen and the best were those for nitrogen.

The problem becomes one of adjusting the experimental data to achieve perfect balances without losing their general representation of the gasification process. Since merely averaging the data is not sufficient, special techniques need to be developed; these are the primary and secondary balance adjustments as described in the following sections.

Primary Balance Adjustments

The strategy of the primary adjustment is to make small adjustments in the rates of the various input and output streams to achieve the best overall and

elemental material balance closure possible, without altering the compositions of any stream. This is based on the assumption that the material imbalances were primarily a result of the errors in the measurement or indirect determination of the stream rates. There were two input streams, air (with moisture) and wood chips (with moisture), and four output streams, dry gas, water, char and tar. There were four elemental species C, H, O and N. Sulfur was neglected due to its small amount in biomass.

Since the amounts of tar and char were very small compared to the other streams, they were not altered in the primary adjustment. Efforts were directed at adjusting the four remaining streams to achieve the best overall results for the five closures. This is a multi-variable multi-objective optimization problem. To solve it, priority must be assigned to each of the five closures. Priorities were chosen to be, in descending order, the overall, C, H, O and N closures. This order was determined according to their relative importance in the subsequent thermodynamic analysis. The problem was then solved using sequential search techniques with the aid of computer.

Table 4 presents the extent of the adjustment on each stream as a percentage of each stream rate. The maximum adjustment was 17%. Table 5 presents the closures for the 8 runs after the first adjustment. As can be seen, all the overall closures are 1.00 (3 digit accuracy).

Table 6 presents the mass balance data (on a dry ash free basis) for the 8 runs after the first adjustment, normalized such that the wood chip rates are all 100. Note that all of the water terms have been combined into a single term, the net water output. The means and standard deviations for the 8 runs are also given at the bottom of the table.

Secondary Balance Adjustments

As can be seen from Table 5, although the closures after the primary adjustment improved, they still lacked perfect closure. Therefore, a secondary adjustment was conducted. The secondary adjustment scheme was developed based on the following considerations:

- (1) As many compositions as possible of the six streams were to remain unchanged. However, all of them can not remain fixed. The composition of char was chosen to be open for adjustment, because (a) the compositions of air and water are fixed, (b) the compositions of wood chips and dry gas were measured and are preferred to remain fixed due to their importance in the thermodynamic analysis, and (c) variation of the composition of tar is not effective, because the amount of tar is too small.
- (2) The amount of dry gas relative to of wood chips was fixed due to its importance in the thermodynamic analysis.

Therefore, the amount of air, the amount and composition of char, and the net water output remained to be adjusted.

The secondary adjustments were based on the means for the 8 runs, as presented in Table 6. The first and final stages of this adjustment are presented in Table 7. In the first stage, the amount of dry air was first slightly adjusted to give a complete overall material balance closure (since the closure 1.00 obtained from the primary adjustment was accurate only up to 3 digits). Then each of the first 5 streams were broken down into their elemental constituents C, H, O, and N. Next, the elemental constituents of char were determined by difference to force perfect elemental balances. However, as can be seen from Table 7, the resulting char composition is not

realistic with respect to oxygen and nitrogen.

From the elemental analysis of the char presented in Table 3, the oxygen and nitrogen contents of char should be very small or negligible. Therefore, the assumption of zero oxygen and nitrogen contents for char was adopted. The nitrogen content of the char was then forced to zero by subtracting the same amount from the nitrogen content of the air and accordingly adjusting the oxygen content of the air to maintain the composition of the air. Next, the oxygen content of the char was forced to zero by adjusting the oxygen content of the net water output to compensate for the change of the oxygen in the air and char. Then, the hydrogen content of water was accordingly adjusted to maintain the composition of water and the hydrogen content of char was adjusted to maintain the material balance for hydrogen. The above procedure gave the results presented as the final stage in Table 7.

Converting the above results from a weight basis to a molar basis (using the mean gas composition of the experimental data, see Walawender et al., 1985) and expressing the dry gas in terms of its respective constituents yields the empirical stoichiometric equation for the overall gasification presented in Table 8. Note that the structurally complicated materials, wood chips, tar, and char, all have an arbitrary molecular weight of 100.

THERMODYNAMIC ANALYSIS

The purpose of this analysis was to evaluate the thermodynamic efficiencies for the commercial gasifier. The approach was based on the first and second laws of thermodynamics. These efficiencies can be used to identify the sources of inefficiency in the process.

Background

The thermodynamic analysis was based on the works of Ishimi et al.(1982) and Shieh and Fan (1982). The thermodynamic background for the analysis to follow will be briefly reviewed.

Reference State

The dead state (see, e.g., Shieh and Fan, 1982) is chosen as the reference state for the evaluation of enthalpy and available energy. This is a state in which the materials (or system) can neither exchange energy with nor perform work on the surroundings; in other words, this is a state in which materials are in equilibrium with their environment. For practical cases, the dead state of a material will assume the most stable state of that material in the surroundings. Table 9 presents the datum level materials, datum level concentrations, specific chemical enthalpies and specific chemical exergies of those materials involved in the present gasification process (see, e.g., Shieh and Fan, 1982). The temperature and pressure of the dead state used in this study are 298.15°K and 1 atm, respectively.

Energy Balance

Applying the first law of thermodynamics to the system shown in Fig. 1 and assuming that potential energy, kinetic energy, etc. are negligible, we have

$$\sum_i \sum_k \bar{\beta}_k n_k)_i + Q_0 + W = \sum_e \sum_k \bar{\beta}_k n_k)_e \quad (1)$$

where Q_0 = heat interchange between the system and surroundings,

W = work supplied by the surroundings,

$\bar{\beta}$ = partial molar enthalpy relative to the dead state. Here, $\bar{\beta}$ is defined as (see, e.g. Szargut and Pelola, 1965; Fan and Shieh, 1980)

$$\beta = H - H_0 \quad (2)$$

where H_0 is the partial molar enthalpy in the dead state. Equation (2) can be further expanded to facilitate its evaluation as follows

$$\begin{aligned} \beta &= H - H^0 + H^0 - H_0 \\ &= (H^0 - H_0) + (H - H^0) \\ &= \beta^0 + \int_{T_0}^T (\partial H / \partial T)_P dT + \int_{P_0}^P (\partial H / \partial P)_T dP \\ &= \beta^0 + \int_{T_0}^T C_P dT + \int_{P_0}^P (V - T(\partial V / \partial T)_P) dP \end{aligned} \quad (3)$$

Here, the superscript 0 refers to the standard state and the subscript 0 to the dead state.

When some of the exit streams are discarded to the environment as wastes, equation (1) can be conveniently expressed as

$$\sum H_i = \sum H_{e,u} + \sum H_{e,d} + \sum H_c \quad (4)$$

Here H_i is the total energy input in form of enthalpy and work, $H_{e,u}$ and $H_{e,d}$, are the usable and discarded portions of the energy output in form of enthalpy and work respectively, and H_c represents the energy consumption consisting of Q_0 .

Available Energy Balance

The available energy (or exergy) of a material refers to the maximum work that the material, when brought to its dead state, can do on the surroundings. Making an available energy balance on the system of Figure 1 gives

$$\sum_i \sum_k \bar{e}_{knk})_i + W = \sum_e \sum_k \bar{e}_{knk})_e + T_0 \sigma \quad (5)$$

where σ is the entropy created in the system. The partial molar exergy, $\bar{e}$, is defined as (see, e.g., Gaggioli, 1961; Szargut and Petela, 1965; Rieckert, 1974; Fan and Shieh, 1980)

$$\bar{s} = (\bar{h} - \bar{h}_0) - T_0(\bar{s} - \bar{s}_0) \quad (6)$$

where $\bar{s}$ is the partial molar entropy. This equation can be further expanded to facilitate practical application as follows

$$\begin{aligned} \bar{s} &= \bar{s} - T_0(\bar{s} - \bar{s}_0) \\ &= (\bar{s}^0 - T_0(\bar{s}^0 - \bar{s}_0)) + \int_{T_0}^T \bar{c}_p (1 - T_0/T) dT \\ &\quad + \int_{P_0}^P (\bar{v} - (T - T_0)(\partial \bar{v} / \partial T)_P) dP \\ &= \bar{s}^0 + \int_{T_0}^T \bar{c}_p (1 - T_0/T) dT + \int_{P_0}^P (\bar{v} - (T - T_0)(\partial \bar{v} / \partial T)_P) dP \end{aligned} \quad (7)$$

where $\bar{s}^0 = \bar{s}^0 - T_0(\bar{s}^0 - \bar{s}_0)$ is named partial molar chemical exergy (see, e.g. Fan and Shieh, 1982).

When some of the exit streams are discarded as wastes, equation (5) can be conveniently expressed as

$$\sum A_i = \sum A_{e,n} + \sum A_{e,d} + \sum A_{dis} \quad (8)$$

where A_i is the total available energy input, $A_{e,n}$ and $A_{e,d}$ are the usable and discarded parts of the available energy output and, A_{dis} is the available energy dissipation containing $T_0\sigma$.

Thermodynamic Efficiencies

There are numerous ways to define the thermodynamic efficiency of a process. Only the first law efficiency, ξ_1 , and second law efficiency, ξ_2 , will be considered in this analysis.

In terms of equation (4), the first law efficiency can be expressed as

$$\xi_1 = \sum H_{e,u} / \sum H_i \quad (9)$$

This efficiency represents the ratio of the enthalpy stored in the usable output to the total energy input (including enthalpy and work) per unit input

mass. In terms of equation (8), the second law efficiency can be expressed as

$$\xi_2 = \frac{\sum A_{e,u}}{\sum A_1} \quad (10)$$

This efficiency represents the ratio of the available energy stored in the usable output to the total available energy input per unit input mass. These two efficiencies are concerned mainly the conservation of enthalpy and available energy, respectively. Variations are possible depending on what is considered to be a usable output.

Computations

The commercial gasifier under consideration is shown in Figure 2. There were two input streams, chips (with moisture and ash) and air (with moisture), and four output streams, dry gas, vapor, tar and char (with ash). Equations (3) and (7) were used to evaluate the enthalpy and exergy for each stream, respectively. Equations (4) and (8) were used to evaluate the first law and second law efficiencies for the process, respectively. All the streams were assumed to enter the system at 298.15°K. The moisture content of wood chips was treated as liquid water, ignoring the interaction between the moisture and wood chips. The amount of ash was small compared to wood chips (less than 1%, see Table 3). The ash was treated as an independent inert material which went through the system and carried away only small amount of sensible heat. The binding energy between ash and wood chips or char was ignored.

Work was done on the system by a motor which powered the gas fan, the rotating grate and char air lock. Compared to the energy input from the chips, the mechanical work was not significant and was neglected in the computation.

A method for the evaluation of β° (specific chemical enthalpy) and ϵ° (specific chemical exergy) of structurally complicated materials from their elemental composition has been described by Shieh and Fan (1982). The method uses the following formulas.

$$\beta^\circ = (1 + 0.15[\text{O}]) (7837.667[\text{C}] + 33888.889([\text{H}] - [\text{O}]/8)) \quad (11)$$

and

$$\begin{aligned} \epsilon^\circ = & 8177.79[\text{C}] + 5.25[\text{N}] + 27892.63[\text{H}] - 3173.66[\text{O}] \\ & + 0.15[\text{O}](7837.667[\text{C}] + 33888.889[\text{H}] - 4236.1[\text{O}]) \end{aligned} \quad (12)$$

where [] represents the elemental composition of the complex substance in weight fraction. The units for both β° and ϵ° are kcal/kg. Equations (11) and (12) were used to evaluate β° and ϵ° for the wood chips, char and tar on a dry ash free basis.

The system was open to the atmosphere. The pressure variation through the system was negligible and, therefore, the pressure of the system was assumed to be 1 atm. Furthermore, ideal gas behavior was assumed for all gas phases. Therefore, all of the partial molar terms involved in the evaluation of the enthalpy and exergy of gas mixtures reduced to ordinary molar terms. The pressure dependent term in equation (3) vanished due to the ideal gas assumption. The pressure dependent term in equation (7) was small compared to other terms and therefore was neglected. Consequently, all the pressure dependent terms were dropped in the evaluation of enthalpies and exergies.

To evaluate the temperature dependent terms, as needed in Equations (3) and (7), the heat capacities of the materials as a function of temperature must be known. Heat capacities of the form

$$C_p = a + bT + cT^2 + dT^3 \quad (13)$$

were used for all the gases, H_2 , C_2H_4 , CO_2 , C_2H_6 , C_3H_8 , N_2 , CH_4 , CO and $H_2O(g)$. The coefficient constants are available in the literature (e.g., Reid, et al., 1976). The sensible heat of crude oil was used to approximate that of tar, due to the lack of better information. The heat capacity of char was approximated by graphite (C). The heat capacity of ash was approximated by silicone dioxide.

The crude product stream was composed of dry gas, water, tar and char. Sometimes, the crude product can be directly used for some purpose, such as heating. In other cases, some of its constituents must be removed before use. In principle, the four constituents of the crude product can be separated. Therefore, in evaluating the efficiencies of the process, four operating modes were considered as follows.

Mode 1: All the output species are recovered as usable products.

Mode 2: Char is discarded and the rest of the constituents are used.

Mode 3: Char and tar are discarded and the rest of the constituents are used.

Mode 4: Only dry cooled gas is recovered as usable product.

The temperature of the crude product was observed to range from 644°K to 700°K; therefore, efficiencies at 3 temperatures, 298°K, 644°K and 700°K were determined. The output temperature of 298°K represents the cold output.

Tables 10 and 11 respectively summarize the enthalpies and exergies for various constituents of the input and output streams. The enthalpies and exergies of each constituent are reported in two parts; one is the chemical energy related portion, β° and ϵ° , and the other is the sensible heat related portion, $\Delta\beta_t$ and $\Delta\epsilon_t$. The latter is evaluated for each output at two temperatures, 644 and 700 °K. Also included in the tables are the percentages of input energy stored in the various output constituents at three different

temperatures. The evaluation for char has been conducted using two elemental compositions; one is from the empirical stoichiometry and the other, from the data in Table 3. The results show that the important energy related constituents in the output are, in descending order, CO, H₂, CH₄ and char.

Table 12 summarizes the first law and second law efficiencies for the four operating modes at the three output temperatures. The char composition used for the evaluation of these efficiencies was based on the data in Table 3. The highest first and second law efficiencies were 89.5% and 62%, respectively, in Mode 1 with an output temperature of 700°K. The lowest first and second law efficiencies were 72% and 53%, respectively, in Mode 4 with an output temperature of 298°K (cold dry gas).

DISCUSSION

Material Balance Analysis

Walawender et al. (1985) reported means and standard deviations for the air input, dry gas output, net water output and char output in kg per kg dry chips (DC) for the 8 runs as 1.62, 0.13; 2.46, 0.19; 0.13, 0.054; and 0.032, 0.0061, respectively. The empirical stoichiometry derived in this study resulted in air input, dry gas output, net water output and char output in kg per kg DC of 1.69, 2.60, 0.056 and 0.036 respectively (see the bottom line of Table 7). Comparison the derived stoichiometry with the experimental data shows that the air input, dry gas output and char output were within one standard deviation of the mean of the experimental data. The net water output was slightly more

than one standard deviation from the mean of the experimental data. Therefore, the derived stoichiometric formula was able to represent the gasification process in terms of the relative amounts of each stream.

All the compositions of various streams, except for char, remained unchanged in formulating the stoichiometric equation. Note that the composition of the tar was arbitrarily assigned, because the amount of tar was negligible. However, since the char was selected as the victim to close the material balance, its composition has been distorted. The molar hydrogen to carbon ratio of the char was determined to be 6.3, which is higher than that of methane. The distortion of the composition of char is the tradeoff for the complete material balance closure. The high hydrogen content of char resulted from the low elemental closures for hydrogen in the original data as shown in Table 2. Consequently, this equation is primitive and needs to be refined when more accurate material balance data become available.

Thermodynamic Analysis

The evaluation of the first and second law efficiencies were based on the stoichiometric formula derived in the material balance analysis. However, there was one adjustment. The composition of char was based on the experimental data to provide a more realistic measure of the gasifier efficiency when char was considered to be a usable product.

The highest first law efficiency, 89.5%, implied that the heat loss of the system to its surroundings is 10.5% of the input enthalpy. The available energy loss due to this heat loss is about 7%. This dissipation was evaluated by assuming that the heat was dissipated from the system at a uniform

temperature 900°K to its surroundings at 298°K. Therefore, the use of more insulation to reduce the heat loss is important, especially in terms of available energy.

The lowest first law efficiency, 72%, corresponds to the cold gas efficiency; this is the same as that reported by Walawender et al. (1985). The lowest second law efficiency, 53%, represents the second law cold gas efficiency. This indicates that when cold product gas is the only usable product, 28% of the enthalpy and 47% of the available energy are lost in the process.

The highest second law efficiency, 62%, along with the available energy dissipation due to the heat loss, 7%, implies that 31% of the input available energy is dissipated in the system due to the various irreversibilities of the gasification process. This means that 31% of the input available energy is dissipated in the system due to the nature of the process. This dissipation can not be recovered.

The variation in the output temperature shows that the sensible heat accounts for 7 to 8% of the enthalpy and only 2 to 3% of the exergy for all 4 operating modes. This indicates that the sensible heat is more valuable in terms of enthalpy than in terms of exergy.

Comparing the efficiencies in the different operating modes shows that, for all output temperatures, (1) discarding the char will lower both efficiencies by approximately 6%, (2) discarding the tar will lower both efficiencies by less than 1%, and (3) discarding steam will lower the first law efficiencies by approximately 3% and has negligible influence on the second law efficiencies. These results indicate that the utilization of the energy (mainly the chemical energy) stored in char can improve the efficiency of this

process significantly.

SUMMARY

An empirical stoichiometry for wood chip gasification in moving bed downdraft gasifier has been derived. The equation was based on an analysis of experimental overall and elemental balance data. However, the char composition in the stoichiometry was distorted.

A thermodynamic analysis was applied to a commercial moving bed downdraft gasifier. This analysis was based on the empirical stoichiometry. The first law and second law thermodynamic efficiencies were evaluated for the process for four different operating modes at three different output temperatures. The highest first law and second law thermodynamic efficiencies were evaluated to be 89.5% and 63% respectively. The lowest first law and second law were evaluated to be 72% and 53% respectively, representing the cold dry gas output. The heat loss from the system to its surroundings was estimated to be 10.5% of the input energy, and 7% of the input available energy was lost due to this heat loss. The available energy dissipation in the system, due to the various irreversibilities of the gasification process, was evaluated to be 31% of the input available energy. This dissipation reflects the nature of the process.

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Table 1. Material Balance Summary

Run	Inputs (kg/hr)			Outputs (kg/hr)						Overall Closure(%)
	Wet chips	Dry air	H ₂ O	Total	Dry gas	Char	Tar	H ₂ O	Total	
98	32.0	43.1	0.5	75.6	66.3	0.9	0.09	7.4	74.7	98.9
910	32.0	45.2	0.5	77.7	68.0	0.9	0.14	7.4	76.4	98.3
920	35.7	62.1	0.4	98.2	92.9	1.4	0.09	7.1	101.5	103.4
101	52.4	76.9	0.8	130.1	116.9	1.7	0.09	13.3	132.0	101.4
929	52.7	74.0	1.0	127.7	113.0	1.8	0.14	12.2	127.1	99.5
106	58.1	76.8	0.5	135.4	117.0	1.2	0.09	10.8	129.1	95.4
1119	89.1	112.1	1.0	202.2	173.8	3.0	0.27	19.7	196.8	97.3
1117	96.2	140.4	0.7	237.3	218.9	2.5	0.18	22.9	244.5	103.0
1221	140.5	202.1	1.1	343.7	302.1	4.1	0.54	43.3	350.0	101.8

Table 2. Mass Balance Closures* of Original Data

Run	C	H	O	N	Overall
98	0.96	0.86	1.01	0.99	0.99
910	0.96	0.85	1.00	0.99	0.98
920	1.06	0.99	1.09	0.99	1.03
929	0.92	0.93	1.05	0.99	1.00
106	0.82	0.84	0.99	0.99	0.95
1119	0.94	0.85	0.98	0.99	0.97
1117	1.02	0.99	1.09	0.99	1.03
1221	0.93	1.02	1.10	0.99	1.02

*Closure is defined as the mass ratio of output to input.

Table 3. Elemental Compositions*

		C	H	O	N	Ash
Wood chips**	mean	48.11	6.05	44.97	0.13	0.74
	σ	0.44	0.09	0.51	0.06	0.21
Dry air		-	-	23.30	76.70	-
Char***	mean	75.83	0.89	4.74	0.06	19.48
	σ	8.15	0.08	-	0.05	-
Water		-	11.11	88.89	-	-

*Weight %.

**9 samples; dry basis.

***8 samples; dry basis.

Table 4. Summary of the Primary Adjustment*

Run	Streams adjusted (%)			H ₂ O**
	Air(wet)	Chips(wet)	Dry gas	
98	0	-10	- 5	15
910	0	- 5	0	0
920	4	5	0	15
929	4	- 8	0	- 5
106	4	-17	0	- 5
1119	0	- 7	0	0
1117	4	2	0	5
1221	4	- 5	0	-15

*Adjustment is expressed by the percentages by which the rates of the corresponding streams were adjusted.

**This is the vapor in the outlet gas stream.

Table 5. Mass balance closures* After Primary Adjustment

Run	C	H	O	N	Overall
98	1.01	0.98	1.07	0.94	1.00
910	1.00	0.90	1.02	0.99	1.00
920	1.00	0.99	1.06	0.96	1.00
929	0.99	0.98	1.06	0.96	1.00
106	0.99	0.99	1.07	0.96	1.00
1119	1.00	0.91	1.02	0.99	1.00
1117	0.99	0.98	1.07	0.96	1.00
1221	0.97	0.99	1.06	0.96	1.00

*Closure is defined as the mass ratio of input to output.

Table 6. Summary of Mass Balance* After Primary
Adjustment on Dry Ash Free (DAF) Basis.

Run	Dry chips + Dry Air	=	Dry gas + H ₂ O	+ Tar	+ Char	
98	100.00	181.41	265.32	14.05	0.38	2.41
910	100.00	179.67	270.23	7.41	0.54	2.61
920	100.00	184.34	265.42	15.88	0.26	3.20
929	100.00	172.86	273.90	15.68	0.31	3.20
106	100.00	178.65	261.88	14.69	0.20	1.84
1119	100.00	163.21	253.05	7.74	0.40	3.18
1117	100.00	168.80	252.98	14.21	0.21	2.11
1221	100.00	178.12	256.04	18.24	0.46	2.29
mean	100.00	175.88	259.85	13.49	0.35	2.61
σ	—	7.06	6.72	3.88	0.12	0.54

*The values were normalized such that all chips are 100.

Table 7. First and Final Stages of the
Secondary Adjustments.

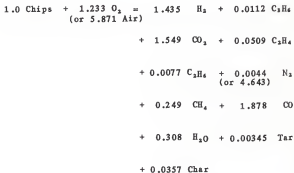
First stage

Element	Chips	+ Air	=	Gas	+ H ₂ O	+ Tar	+ Char
C	48.477	0.000		45.965	0.000	0.168	2.344
H	6.094	0.000		4.215	1.499	0.021	0.359
O	45.306	41.076		79.622	11.991	0.156	-5.387
N	0.123	135.215		130.050	0.000	0.000	5.288
Sum	100.000	176.291		259.852	13.490	0.345	2.604
Total input = 276.291				Total output = 276.291			

Final stage

Element	Chips	+ Air	=	Gas	+ H ₂ O	+ Tar	+ Char
C	48.477	0.000		45.965	0.000	0.168	2.344
H	6.094	0.000		4.215	0.625	0.021	1.233
O	45.306	39.470		79.622	4.998	0.156	0.000
N	0.123	129.927		130.050	0.000	0.000	0.000
Sum	100.000	169.397		259.852	5.553	0.345	3.577
Total input = 269.397				Total output = 269.397			

Table 8. An Empirical Equation of Gasification of
Wood chips in a Moving Bed Gasifier.



where Chips, Tar and Char all have arbitrary molecular
weight of 100 and their empirical formulas are



Table 9. Specific Chemical Enthalpy, Specific Chemical Exergy,
Datum Level Material and Datum Level Concentration.

	Datum level material	Datum level concentration (molar frac.)	β^{**} (kcal/mol)	ϵ^{***} (kcal/mol)
C(s)	CO ₂ (g)	0.000302	94.052	98.131
CO(g)	'	'	67.630	65.790
CO ₂ (g)	'	'	0.0	4.802
CH ₄ (g)	CO ₂ (g), H ₂ O(l)	0.000302, 1	212.800	152.380
C ₂ H ₄ (g)	'	'	337.240	278.950
C ₂ H ₆ (g)	'	'	372.820	287.990
C ₁ H ₄ (g)	'	'	491.990	408.990
H ₂ (g)	H ₂ O(l)	1	68.320	33.197
H ₂ O(l)	'	'	0.0	0.0
H ₂ O(g)	'	'	10.511	0.0
N ₂ (g)	N ₂ (g)	0.78	0.0	0.147
O ₂ (g)	O ₂ (g)	0.2096	0.0	0.926
Air	N ₂ (g), O ₂ (g)	0.78, 0.2096	0.0	0.0

*Specific chemical enthalpy.

**Specific chemical exergy.

Table 10. Summary of Enthalpy Evaluation

	n (mole)	β^0 ----- (kcal/mol)	$\Delta\beta_{f,0}^{0,0}$ ----- (kcal/mol)	$\Delta\beta_{f,0}^{0,0}$ ----- (kcal)	% of input	$n\Delta\beta_{f,0}^{0,0}$ (kcal)	% of input	$n\Delta\beta_{f,0}^{0,0}$ (kcal)	% of input
Input: Chips	1.0000	421.360		421.360					
Air	5.8715	0.000		0.000					
H ₂ O(chips)	0.6866	0.000		0.000					
H ₂ O(air)	0.0201	10.511		0.211					
Ash(chips)	0.0080	0.000		0.000					
Total				421.571					
Output: H ₂	1.4348	68.320	2.417	2.808	23.25	101.493	24.075	102.054	24.208
C ₂ H ₄	0.0112	491.990	7.439	8.973	1.31	5.594	1.327	5.611	1.331
CO ₂	1.5491	0.000	3.593	4.244	0.00	5.566	1.320	6.574	1.559
C ₂ H ₆	0.0509	337.240	4.988	6.006	4.07	17.471	4.132	17.419	4.144
C ₂ H ₄	0.0077	372.820	6.144	7.423	0.68	2.918	0.692	2.928	0.695
N ₂	4.6425	0.000	2.447	2.853	0.00	11.361	2.895	13.246	3.142
CH ₄	0.2487	212.800	3.743	4.489	12.55	53.854	12.775	54.040	12.819
CO	1.8782	67.630	2.462	2.874	30.13	131.648	31.228	132.420	31.411
H ₂ O	1.0147	10.511	2.909	3.403	2.53	13.617	3.230	14.119	3.349
Tar	0.0034	465.911	19.830	24.020	0.38	1.676	0.398	1.690	0.401
Char** (a)	0.0357	1684.618	9.508	11.439	14.23	60.362	14.318	60.431	14.335
(b)		747.848			6.32	26.985	6.401	27.053	6.417
Ash	0.0080	0.000	7.908	9.406	0.00	0.063	0.015	0.075	0.018
Total** (a)					89.15	405.570	96.205	410.658	97.411
(b)					81.23	372.193	88.287	377.281	89.493

* $\Delta\beta_t$ represents $(\beta_t - \beta^0)$, where t is the temperature in $^{\circ}\text{K}$.

**Two compositions of char were used for evaluating its specific chemical enthalpy; for (a) the composition from the table 8 was used, for (b) the composition of char from Table 3 was used.

Table 11. Summary of Energy Evaluation

	n (mole)	Δe_{s-t} (kcal/mol)	Δe_{t-s} (kcal)	% of input	Δe_{s-t} (kcal)	% of input	Δe_{t-s} (kcal)	% of input
Input :Chips	1.0000	449.441			449.441			
Air	5.8715	0.000			0.000			
H ₂ O(chips)	0.6866	0.000			0.000			
H ₂ O(air)	0.0201	0.000			0.000			
Ash(chips)	0.0080	0.000			0.000			
Total					449.441			
Output:H ₂	1.4348	33.197	0.815	1.032	47.631	10.60	48.800	10.858
C ₂ H ₄	0.0112	408.990	2.673	3.527	4.581	1.02	4.611	1.026
CO ₂	1.5491	4.802	1.248	1.611	7.439	1.66	9.373	2.085
C ₂ H ₆	0.0509	278.950	1.787	2.354	14.199	3.16	14.290	3.179
C ₂ H ₂	0.0077	287.990	2.211	2.923	2.217	0.49	2.234	0.497
N ₂	4.6425	0.000	0.827	1.053	0.000	0.00	3.841	0.855
CH ₄	0.2487	152.380	1.326	1.742	37.897	8.43	38.227	8.505
CO	1.8782	65.790	0.834	1.063	123.567	27.49	125.133	27.842
H ₂ O	1.0147	0.000	0.990	1.265	0.000	0.00	1.005	0.224
H ₂	0.0034	491.484	28.480	33.236	1.696	0.38	1.794	0.399
Char** (a)	0.0357	1499.518	13.656	15.827	53.428	11.88	53.914	11.996
(b)		779.122			27.760	6.18	28.247	6.285
Ash	0.0080	0.000	2.782	3.615	0.000	0.00	0.022	0.005
Total** (a)					292.654	65.12	303.243	67.471
(b)					266.986	59.40	277.575	61.760
								280.454
								62.401

* Δe_t represents ($e_t - e_s$), where t is the temperature in °K.

**Two compositions of char were used for evaluating its specific chemical energy; for (a) the composition from the Table 8 was used, for (b) the composition of char from Table 3 was used.

Table 12. Summary of First Law and Second Law Efficiencies

First law efficiency			Unit: %		
Mode	Discarded output	Output temperature(*K)			
		298	644	700	
1	None	81.24	88.30	89.51	
2	Char(ash)	74.92	81.89	83.08	
3	Char(ash),tar	74.54	81.49	82.67	
4	Char(ash),tar,H ₂ O	72.00	78.24	79.31	
Second law efficiency					
Mode	Discarded output	Output temperature(*K)			
		298	644	700	
1	None	59.40	61.76	62.40	
2	Char(ash)	53.23	55.47	56.09	
3	Char(ash),tar	52.85	55.07	55.69	
4	Char(ash),tar,H ₂ O	52.85	54.85	55.40	

Note the evaluation of enthalpy and exergy for char was based on the measured composition of char (presented in Table 3).

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CHAPTER 5 GASIFICATION OF WOOD PELLETS AND CHIPS IN A DOWNDRAFT
GASIFIER

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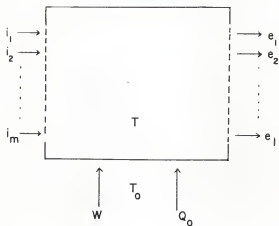


Figure 1. Schematic Diagram of an Open Flow System.

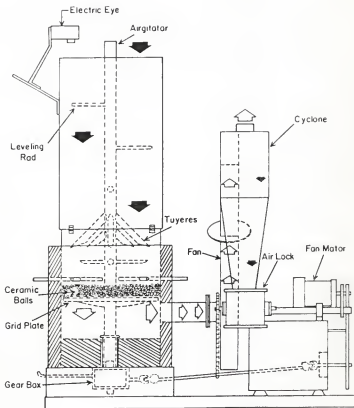


Figure 2. Schematic of the Gasifier.

CHAPTER 5

GASIFICATION OF WOOD PELLETS AND CHIPS
IN A DOWNDRAFT GASIFIER

INTRODUCTION

Gasification is one means for converting wood or any biomass into a gaseous fuel which is more versatile than the original biomass. Currently, wood provides about 2% of our total energy needs and could contribute up to 8% within the next decade (Zerbe, 1981). The annual harvest of woody biomass amounts to about 1.4 billion tons in the U.S. alone. Over 700 million tons of this material is not used because it is not of the right species, size, fiber length, fiber morphology etc. (Goldstein, 1978). Branchwood is a prime example of this type of wood waste. Thus the potential for the utilization of wood to ensure a continuous supply of fuels and chemicals is significant.

Some advantages of wood as a gasification feedstock over solid fossil fuels such as coal and oil shale are its higher reactivity, its lower sulphur content, and its lower ash content. Most important is the fact that it is renewable. Its disadvantages stem from its lower heating value, lower bulk density and its higher moisture content which makes handling facilities more expensive. In this regard, beneficiation processes (see, e.g., Bain, 1980) are available for improving the properties of biomass, such as drying and densification. Wood pellets are a prime example of improved biomass.

Various types of reactors have been used to gasify biomass; they include fixed beds, entrained beds, moving beds, rotary kilns, and fluidized beds. Fixed bed and moving bed gasification has been practiced for a long time and a few commercial units are available. The other technologies are relatively new, and are basically geared to large-scale operation; only a very few systems have been commercialized. Reed et al. (1980) summarized the state of gasifier research and the status of various gasifier manufacturers.

Downdraft gasifiers for the production of low BTU gas from biomass are generally mechanically simple and relatively easy to operate with suitable feed materials. The product gas has a negligible tar content. This feature makes them superior to updraft gasifiers which produce a gas with a high tar content (about 20%) when biomass is used as the feedstock. The clean gas can be used for different applications, such as heating, and fueling boilers or engines.

Moving bed gasifiers have been used for the production of low BTU gas since the early nineteenth century. Sweden (Reed and Jantzen, 1979) and other countries used them to power automobiles during World War II. However, their performance under various operating conditions has not been well understood. In this regard, numerous research efforts are currently underway to improve our understanding and to develop models to facilitate gasifier design. Despite the limited knowledge, a large number of gasifiers have been constructed for research and/or commercial use. The Buck Roger's TM gasifier is one recently commercialized downdraft gasifier which has undergone some design changes since its invention in 1982. The gasifier used in this study is a variation of its predecessor used in our previous work (Walawender et al., 1985).

The objective of this work was to determine the performance of the gasifier under different operating conditions through complete material balance data for the gasifier. The operating parameters studied included the rotation speed of the grate, type of feed, and bed supports. Two feedstocks, wood chips and wood pellets, were gasified and three different bed supports were investigated at grate rotation speeds ranging from 3 to 14 min/rev.

EXPERIMENTAL FACILITIES AND PROCEDURE

Gasifier Description

The Current gasifier

Figure 1 presents a schematic diagram of the gasifier used in this study. It has a diameter of 0.6m and is open at the top. There is no throat or choke plate in the unit. A fan located downstream from the gasifier draws air through the bed as well as the tuyeres. A butterfly valve inserted in the fan outlet serves as throttle for the coarse control of gas throughput. A feed bin and a screw feeder (not shown in Figure 1) supply feed to the top of the gasifier each time the bed height drops to a set level. Feeding is controlled by an electric eye, mounted at the side of the reactor. A plate (on the top of the reactor) with a bearing is used to hold the upper end of the 'Airgitator' in place. A chute is seated on the top plate to receive the feed delivered by the screw feeder. The 'Airgitator' is driven by a hydraulic motor with a hydraulic fluid flow splitter for the control of rotation speed. It rotates at speeds from 3 to 14 minutes per revolution. The 'Airgitator' has several functions, listed below:

- (1) It provides part of air intake, through the tuyeres.
- (2) It levels the feed at the top of the bed.
- (3) It slowly mixes material throughout the gasifier.

Gas and char flow out of the gasifier through the grid plate which is attached to the 'Airgitator' shaft. They are drawn out of the base of the gasifier by the fan which blows them through the cyclone where the char is

separated from the gas. The gas stream is then sent to a flare stack (not shown in Figure 1) where it is incinerated.

Comparison With the Previous Gasifier

The current gasifier is basically similar to its predecessor (used in our previous study (Walawender et al., 1985)) in most aspects, except for a few design changes which are detailed as follows. The current gasifier allows the grate rotation speed to be controlled independent of the fan speed, whereas its predecessor had the two speeds proportional to each other through a series of speed reducing pulleys and gears. The rotation speed of the fan in the current gasifier is fixed (or more precisely, the applied voltage to the fan motor driving the fan is fixed), whereas the rotation speed of the fan in the previous gasifier was electronically controllable. Therefore, a butterfly valve is used at the outlet of the fan as a throttle for the coarse control of gas throughput in the current system.

Operating Procedure

The operating procedure for the gasifier is detailed below.

1. Start-up. A portable propane burner was used to preheat the system for about 15 minutes. The temperature in the empty gasifier was about 600°K at the end of the preheat period. During the preheat period, the fan was started and its throughput gradually increased from low to maximum. The fan was turned down to low throughput at the end of the preheat period. Propane was then introduced into the flare stack and ignited. Next, the feeder was manually operated to fill the gasifier to build a shallow bed about 4 inches thick. The bed was then set on fire. After the fire was uniformly spread

over the bed, another layer of feed was introduced. The same procedure repeated until the bed level was just above the tuyeres. The gasifier was then filled to the desired operating level, and the automatic feeding mode was turned on. The grate was rotated very slowly or even stopped when the bed was being developed. The rotation speed was gradually increased up to the desired speed after the bed was established. As the gas production and gas temperature increased, the propane rate into the flare stack was decreased and finally shut down. The fan throttle was finally brought to the wide open position and the gasifier was allowed to attain steady state. This procedure normally required two hours to complete.

2. Gas analysis and condensables measurement. After the completion of step 1, a side drawn of the gas was taken for analysis and condensate determination. The specific procedures will be detailed in the next section. These measurements were conducted for a two hour period.

3. Char measurement. Char was measured by direct collection over a one to two hour intervals in the middle of the run.

4. Tracer gas measurement. Nitrogen tracer gas was introduced into the product gas (at the gas inlet) at a known rate, and the gas analysis and condensate measurement were repeated. The specific procedures are detailed in the next section. These measurements were conducted for one hour.

5. Post tracer measurement. Step 3 was again repeated without the tracer gas for 30 minutes to one hour to confirm the steady state of the system.

6. Shut-down. The supply of feed was stopped and the gasifier continued to consume the feed in the bed. As the bed was depleted, flame began to spread over the surface of the bed. Then, the whole bed reduced to char and the temperature of the bed began to rise rapidly. The fan was shut down when the

temperature of the bed reached 1200°K and the remaining char was allowed to burn slowly. Normally more than four hours were required to completely burn the char.

Measurements

Because of the configuration of the gasifier and the stream flow rates involved, it was not convenient to measure all of the stream rates directly; consequently, indirect methods were employed.

Direct Measurements

(1) Char output rate. Char output was determined by collecting and weighing the char over a one to two hour interval in the middle of the run, usually covering the tracer gas measurement period.

(2) Gas analysis. The composition of gas with and without nitrogen tracer was determined with an Applied Automation on-line process gas chromatograph (GC) which drew a continuous sample from a point downstream from the cyclone. The GC had a cycle time of about 11 minutes and was able to detect the following components in the producer gas: H_2 , CO_2 , CO , CH_4 , O_2 and N_2 along with traces of C_2H_6 , C_2H_4 and C_2H_2 .

(3) Condensables. Condensables consisted of two parts, tar and aqueous. These were determined by taking a side draw from the main gas stream at a rate of about 0.55 m³/hr. This stream also supplied the GC with sample. The gas was first sent through a hot filter which was packed with glass wool to remove heavy tar components. Next the gas was sent through two water cooled condensers in series to remove most of the water. The remaining water removal

was accomplished with two receivers placed in series in an ice bath. The remaining tarry mist was then removed by passing the stream through a tightly packed glass wool filter. Flow through the sample train was maintained with the aid of a 'Gast' compressor which provided suction. The discharge from the compressor was then sent through a wet test meter, followed by a drierite chamber and then to the GC. The wet test meter readings were corrected for temperature, pressure, water of saturation and air, and tracer gas when needed. The aqueous condensates were collected and weighed. The tar was collected, however, it was not measured, because prior experience showed that negligible amount of tar was present (Walawender et al., 1985). The known volume of dry gas flowing through the sample train allowed the amount of condensate to be converted to a mass per unit volume dry gas basis.

(4) Temperatures. Temperatures at various positions in the system were monitored over the course of each run. Six thermocouples were placed in the system to monitor the temperatures at the following positions: just above the the upper tnyeres (drying zone), 0.1 m above the grid plate (one 1 cm from the shaft and one half way between the wall of reactor and shaft), 0.1 m below the grid plate, just before the fan inlet, and at the outlet of cyclone. The temperature data were recorded at 15 to 60 minute intervals.

Indirect Measurements

(1) Dry product gas rate. The dry product gas rate was indirectly determined with the aid of a tracer technique. Nitrogen gas was introduced into the product gas at a known volumetric rate at the fan inlet. From the known tracer injection rate and the nitrogen composition of the gas with and without tracer, the dry gas volumetric flow rate could be readily evaluated by a nitrogen balance. Knowing the molecular weight of the gas allowed the dry gas volumetric rate to be converted to a mass basis.

(2) Dry air and moisture input rates. The dry gas flow rate, coupled with a nitrogen balance between the inlet and outlet of the gasifier, allowed the dry air input rate to be evaluated. This balance assumed that negligible nitrogen was produced in the gasification process. Knowing the temperature and relative humidity of the ambient air over the course of an experiment allowed the moisture input from humid air to be evaluated. Since most experiments were of five hour duration, the moisture evaluation was based on mid-run conditions.

(3) Feed rate. The feed rate for each run was determined by a carbon balance between the inputs and outputs of the gasifier. Among the inputs, only the feedstock contained carbon which was determined by elemental analysis. Carbon existed in two output streams, the product gas and char; the former was determined from the gas analysis and the latter, by elemental analysis. With the rate of gas and char already determined, the feed rate could then be calculated from a carbon balance on the system.

(4) Condensate rate. The directly measured amount of condensate (in mass per unit volume dry gas) and the calculated dry gas flow rate allowed the total condensate output rate to be evaluated.

Operating Parameters

Three operating parameters were investigated in this study, the type of feedstock, bed support and grate rotation speed.

Feedstocks

Two feedstocks were used in this study; they were mixed hardwood chips

(wood chips) and wood pellets (contaminated with about 10% wood chips). The properties of the feedstocks are detailed later.

Bed Supports

The bed was supported by a rotating grate. The grid plate had a diameter of 56 cm with 48 open slots. Each slot had dimensions of 11x52 mm and was rounded at the ends. The open area of the grid plate was estimated to be approximately 12% of the cross section of the gasifier. Three different bed supports were used in turn in this study; the first was with 9 cm bed of ceramic balls (19mm diameter) on the grid plate (W/B), the second, without the balls, but with about half of the open area blocked by broken ceramic balls (W01), and last, an open grid plate (W0).

Grate Rotation

The grate rotation speed was independently controlled. The rotation speed used in this study ranged from 3 to 14 minutes per revolution.

Chemical and Physical Analyses

These analyses included moisture and ash analyses of the feed and char, and elemental analysis of the feed and char. Moisture was determined by drying in an oven for three hours at 380°K. Ash was determined in a muffle furnace by standard ASTM procedure. The elemental analyses of feed and char were conducted with a Perkin-Elmer Model 240b Elemental Analyzer.

Feedstock properties

Table 1 presents various properties of the two feedstocks, wood chips and wood pellets. The wood chips contained a minor portion of bark. The wood pellets were contaminated with about 10% wood chips. The properties presented in Table 1 include the elemental analysis (C,H,O and N), ash content (dry basis), moisture content (wet basis), bulk density, size distribution and higher heating value.

TREATMENT OF DATA

Calculations

The performance of the gasifier was evaluated in terms of a variety of measures determined from the gas composition, material balance data and chemical and physical analyses. Material balance closure was defined as total mass output divided by total mass input. The inputs included air (with moisture) and the wet feedstock. The outputs included dry product gas, condensate, and char. The dry gas composition provides one measure of gasifier performance and from it, the higher heating value of the dry gas was evaluated from the standard heats of combustion of the dry gas components. The gas and char yields provide other measures of performance. These were calculated on the basis of a unit mass of dry feed to eliminate the influence of feed moisture variations. An ash-free basis was not used since the ash contents of the dry chips and pellets were less than 1 %. Char yield was

expressed as the mass of char per unit mass of dry feed (DF) and gas yield was expressed on a volumetric basis as m^3/kg DF. The volume basis used in this work was 289°K and 101.3kPa. Also evaluated were the water output ratio as mass of condensate per unit mass DF and the air-to-feed ratio as mass of dry air per unit mass DF. The total energy output (cold gas) was evaluated in MJ/hr and the energy yield (cold gas) in MJ/kg DF. The cold gas efficiency was evaluated as the ratio of the higher heating value of the dry gas produced from a unit mass DF to the higher heating value of a unit mass DF. This efficiency represents the fraction of the energy content of the feed converted to combustible gas. The mass conversion efficiency was evaluated as the ratio of mass of the dry gas output to the total mass input.

Statistical Analysis

Means and Standard Deviations

Means and standard deviations were evaluated for the various performance measures, including the dry gas higher heating value, volumetric gas yield, water output rate, dry air input, energy yield, mass conversion efficiency, cold gas efficiency, the highest observed temperature (either of the two temperatures monitored 0.1m above the grate) and the dry gas composition. For each performance measure, the statistics were evaluated for all runs and for the runs with closures greater than 95%. For the dry gas composition, the statistics were evaluated for all runs, for runs with wood chips and for runs with wood pellets.

Regression Analyses

For the performance measures which appeared to show distinct trends with changes in the operating parameters or other performance measures, regression analyses using the SAS (Statistical Analysis System) software package were conducted. These analyses included (1) the char yield, cold gas efficiency and air ratio as a function of the grate rotation speed with the type of feed as a parameter, (2) the cold gas efficiency and air ratio as a function of the char yield, and (3) the total energy output as a function of the dry feed rate. The regression models and model parameters are summarized in Table 4, along with the square of the correlation coefficient (R^2) and the probability of falsely rejecting the proposed regression model (prob. F -value $>$ F -statistic in F -test for significance of regression). The parameters within parentheses in the first six models (see Table 4) were determined by a search technique and then the remaining ones were determined by linear regression. The first six models were transformed to logarithmic scales prior to the conduct of linear regression.

RESULTS AND DISCUSSION

Results

A total of 16 runs were completed, and 14 had material balances within 5% of closure, (i.e., 95-105%); the other 2 runs had closures within 6%. Table 2 presents the operating parameters, feed species (with the moisture content of the as-received feed), bed support and grate rotation speed, and various

measures of the gasifier performance. The table is arranged according to the feed species and the type of bed support. Included in the table are the dry feed (DF) rate in kg/hr, the gas higher heating value (HHV), the volumetric gas yield, the char yield, the energy yield MJ per kg DF, the dry air input rate in kg/kg DF, the mass conversion efficiency, the cold gas efficiency and the highest observed temperature. The means and standard deviations for the complete data set and the 14 runs with closures greater than 95% are given at the bottom of the table. The rotation speed of the grate varied from 3 to 13.7 min/rev. The dry feed rate ranged from 18.2 to 74.3 kg/hr. The gas HHV varied from 5.64 to 6.45 MJ/m³ with a mean of 6.07 MJ/m³. The volumetric gas yield varied from 1.35 to 2.20 m³/kg DF with a mean of 1.95 m³/kg DF. The char yield varied from 4.38 to 28.31 kg/kg DF with a mean of 12.51 kg/kg DF. The energy yield varied from 8.33 to 13.64 MJ/kg DF with a mean of 11.80 MJ/kg DF. The total energy output ranged from 216 to 824 MJ/hr. The water output rate ranged from 0.148 to 0.481 kg/kg DF with a mean of 0.216 kg/kg DF. The dry air input rate varied from 0.97 to 1.60 kg/kg DF with a mean of 1.40 kg/kg DF. The mass conversion efficiency varied from 72.1 to 89.7% with a mean of 85.6%. The cold gas efficiency ranged from 49.1 to 80.3% with a mean of 69.5%. The highest observed temperature ranged from 853 to 1172°K with a mean of 1010°K.

Table 3 summarizes the dry gas compositions (mole %) for all the runs. The principal components of the product gas were H₂, CO, CO₂, CH₄ and N₂ with trace amounts of C₂H₄, C₂H₆ and C₂H₂. The means and standard deviations for all runs, for wood chips, and for wood pellets are also provided at the bottom of the table. Although the data show some fluctuation, no distinct trends were observed as the operating parameters varied. However, the means for the hydrogen composition were significantly different for wood chips, 12.2%, and for wood pellets, 16.11%; this was determined by applying the pooled t-test to

determine the significance of difference of the two means, which gave 0.001 for Prob.>|t|.

Figures 2 through 4 present the variations of char yield, cold gas efficiency, and air ratio as a function of the rotation speed of the grate, respectively. The data are presented in 3 groups, (1) for wood chips with the ball bed (4 runs), (2) for wood pellets with the ball bed (2 runs), and (3) for wood pellets without the ball bed (10 runs). For groups 1 and 3, asymptotes were observed as the grate rotation approached 3 min/rev. The regression lines (see Table 4) for groups 1 and 3 are included in the figures. Unfortunately, the sample sizes were too small to statistically test for significant differences between the regression models for groups 1 and 3, although a definite difference appears to exist.

Figures 5 and 6 present the variations of the cold gas efficiency and the air to feed ratio as a function of the char yield, respectively. Although the data show some fluctuation, a linear correlation was determined by the least-squares method. The parameters and statistics are included in Table 4. The two lowest points in Figures 5 and 6 correspond to the two points at 3 min/rev in Figures 2, 3, and 4.

Figure 7 shows the total energy output as a function of the dry feed rate for the present gasifier, as well as for the previous gasifier (Walawender et al., 1985) for comparison. In fitting the present data by the least-squares method, the two data points at 3 min/rev were excluded. As can be seen from the figure, both data sets were well described by straight lines. Table 4 presents the statistics. It is interesting that these two lines almost coincide with each other, considering the large differences in grate rotation speeds in the two gasifiers. The previous gasifier was operated with maximum grate rotation of the order of 20 min/rev.

Discussion

The 16 experimental runs can be classified, according to feed species and the type of bed support, into three groups, (1) wood chips with the ball bed (4 runs), (2) wood pellets with the ball bed (2 runs), (3) wood pellets without the ball bed, including (a) W01 (3 runs) and (b) W0 (7 runs). Since the sample sizes of the three groups are relatively small, a statistical analysis for determining the effect of the operating parameters on the gasifier performance is not possible. However, a rather primitive qualitative assessment can be made.

Comparison of the results for groups 2 and 3 (a and b) reflects the influence of the bed support on the gasifier performance, for wood pellets. Only a few of the performance measures exhibited noticeable trends as the rotation speed of the grate increased. These are shown in Figures 2 through 4. As can be seen from the Figures, the data points for groups 2 and 3, although scattered, show no definite trend from one group to another as the rotation speed of the grate varied. Therefore, it appears the the three different bed supports have no significant influence on the gasifier performance.

Comparison of the data for group 1 and 3 illustrates the influence of feed type on the gasifier performance. As can be seen from Figures 2 through 4, different regression lines described the data for the two groups. The char yield was lower for chips than for pellets, the cold gas efficiency was higher for chips than for pellets, and the air ratio was higher for chips than for pellets.

The shortest average residence time of the solid phase in the active zone

of the gasifier for the 16 completed runs was estimated to be in the order of 20 minutes. The average residence time was evaluated by dividing the length of active zone of the gasifier (approximately 0.3m) by the average solid phase velocity in the active zone. The time required for complete pyrolysis of various types large particles in a moving bed downdraft gasifier was reported by Reed and Markson (1982) to be less than 5 minutes. Therefore, it can be concluded that the residence time of solid phase in this study was sufficient for complete pyrolysis of the solid phase in all the runs. Consequently, the residence time was not responsible for the difference in char yield between wood chips and wood pellets.

The difference in char yield between wood chips and wood pellets is probably due to the differences in the dimensions and physical properties (e.g., density and porosity) of the two feedstocks. It is known that char yield is higher for large particle than for small particle. Chan et al. (1985) reported that the char yields for wood pyrolysis under fire-level radiation were 19% and 24% for the pellets with characteristic dimensions of 0.5 cm and 1.5 cm, respectively. The smallest dimension (representing the characteristic dimension) of the wood chips ranged from 2 to 5 mm, and that of wood pellets ranged from 12 to 14 mm. The dimension ratio of pellets to chips ranged from 3 to 7. Therefore, the char yield of wood pellets can be expected to be higher than wood chips because of their differences in dimensions. In addition, the wood pellets were about twice as dense as the wood chips (see Table 1). Therefore, the particle porosity for pellets can be expected to be smaller than that for wood chips (by a factor of about 1/2). The variation of char yield with the particle density or porosity of the feedstock has not been determined. However, it is intuitively expected that it has a role in determining the char yield.

As the grate rotation speed increased, a sharp increase in char yield and a sharp decrease in the air to feed ratio were observed. The air to feed ratio dropped to about 1.2 and 1.0 for wood chips and wood pellets respectively. Our previous study (Walawender et al., 1985) showed that by maintaining an average air to feed ratio at 1.6, we were able to obtain an average char yield of 3.3% for the gasification of wood chips. Furthermore, Run 08/29 of this study had a char yield of 4.8% with an air to feed ratio of 1.6. These observations suggest that the increase in char yield with increasing grate rotation speed was a result of the decrease in the air to feed ratio. This is supported by the decreasing linear relationship between the air to feed ratio and the char yield shown in Figure 6. Unfortunately, the air to feed ratio for the present gasifier can not be independently controlled. In principle, the air input will be determined by the depth and porosity of the bed and the pressure drop across the bed. The fast rotation of grate might have caused a dramatic change in the bed porosity due to the action of the tuyeres and the pressure distribution in the bed, which in turn resulted in a sharp increase in the char yield. This was suggested by the decrease in the total air input at high grate rotation speed.

The cold gas efficiency was found to decrease linearly with increasing char yield. This relationship along with the decreasing linear relationship between the air to feed ratio and char yield implied a linear relationship between the air to feed ratio and cold gas efficiency. Note that the cold gas efficiency represents a conservative estimate of the actual first law efficiency for the gasifier. In other words, by maintaining an adequate air to feed ratio a high efficiency will result.

Table 5 summarizes the operating conditions and various performance measures for the present study and the previous study (Walawender et

al.,1985). As can be seen from the table, despite the differences in operating conditions used for the two gasifiers, most of the performance measures are essentially the same, including the gas yield, water output, energy yield, cold gas efficiency, mass conversion efficiency and gas composition. However, the average char yield of the present gasifier was three fold larger than that of the previous gasifier. The average air input of the present gasifier, 1.44 kg/kg DF, was also significantly different from that of the previous gasifier, 1.60 kg/kg DF; this is supported by the pooled t-test, which gives 0.005 for Prob.>|t|. Another difference is the dry gas HHV, which was 6.08 MJ/m³ for the present gasifier and 5.51 MJ/m³ for the previous one; the Prob.>|t| for the pooled t-test was 0.001. The energy output rate of the present gasifier was lower than that of the previous one almost by a factor of 1/2, however the throughput was also lower.

CONCLUSIONS

The results of this work show that the performance of the gasifier used in this study is similar to the gasifier used in previous study. From the viewpoint of the total energy output, the gasifier treated wood chips and wood pellets as if they were the same.

The char yield was shown to sharply increase at high rotation speed of the grate and level out at slow rotation speed of the grate. The air to feed ratio and cold gas efficiency were shown to decrease linearly with increasing char yield. The cold gas efficiency was shown to increase linearly with increasing the air input.

The char yield from wood pellets was shown to significantly higher than

that from wood chips. The higher yield was a consequence of the higher characteristic dimension.

The use of three different bed supports was shown to have no significant influence on the performance of the gasifier when comparable operating conditions were used.

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Table 1. Properties of Feedstocks

		Wood chips	Wood pellets
Elemental and	C	47.07	48.59
ash analysis	H	5.84	5.96
(dry wt. %)	O	46.29	44.23
	N	0.35	0.37
	Ash	0.45	0.85
Moisture (wet wt. %)		9.99	6.50
Bulk density (kg/m ³)		236	585
HHV* (MJ/kg)		16.0	17.1
Size distribution (wt. %)			
	1>12.7mm	15.4	1~ 2cm
	12.7>1> 4.7mm	74.1	d=12mm 33
	4.7>1> 0 mm	10.5	13mm 33
			14mm 33

*Higher Heating Value, evaluated by Dulong's formula.

Table 2. Performance Summary

Run	Feedstock	Feed Moisture (% wet basis)	Bed	Rotation of Grate(min/r)	Dry Feed rate(kg/hr)
08/27	Chips	10.30	W/B ¹	5.25	64.2
08/28	Chips	9.99	W/B	4.66	57.9
08/29	Chips	9.67	W/B	3.00	74.3
08/29	Chips	9.67	W/B	9.90	38.8
09/10	pellets ¹	6.71	W/B	4.23	39.8
09/11	pellets	6.11	W/B	6.00	18.2
09/24	pellets	6.49	WO1 ³	13.70	44.5
09/28	pellets	6.49	WO1	10.80	49.5
10/01	pellets	6.49	WO1	7.33	56.9
10/03	pellets	6.74	WO ⁴	12.50	58.8
10/05	pellets	6.42	WO	9.58	56.6
10/05	pellets	6.42	WO	6.13	70.0
10/08	pellets	6.49	WO	9.65	60.8
10/10	pellets	6.49	WO	3.00	46.3
10/12	pellets	6.49	WO	9.83	41.1
10/15	pellets	6.49	WO	10.00	57.0

¹Wood pellets contaminated with about 10% wood chips.

²With ceramic balls bed.

³Without ceramic balls bed but with the grid plate half blocked.

⁴Without ceramic balls bed.

Table 2. Continued

Run	Matl. Balance Closure(%)	Dry Gas HHV (MJ / m ³)	Gas Yield (m ³ / kg DF ^a)	Char Yield (kg/100kg DF)	H ₂ O Out (kg/kg DF)
08/27	100.74	5.911	2.00	9.53	0.303
08/28	103.06	6.064	2.00	9.95	0.359
08/29	106.30	5.945	1.55	21.29	0.481
08/29	104.28	6.190	2.20	4.38	0.354
09/10	100.00 ¹	6.261	1.96	10.75	0.214 ²
09/11	98.38	5.993	1.98	10.64	0.184
09/24	101.72	5.640	2.08	12.22	0.301
09/28	103.94	5.829	2.10	9.90	0.261
10/01	102.01	6.145	1.93	11.71	0.234
10/03	104.12	6.186	1.95	13.12	0.312
10/05	99.56	6.023	2.00	13.14	0.202
10/05	100.00 ¹	6.231	1.89	13.74	0.209 ²
10/08	101.91	5.982	2.18	8.44	0.251
10/10	94.31	6.171	1.35	28.31	0.171
10/12	98.16	6.450	1.87	12.82	0.148
10/15	100.19	6.142	2.07	10.26	0.187
All data					
mean	—	6.071	1.95	12.51	0.261
σ	—	0.193	0.22	5.455	0.0872
>95%					
mean	—	6.075	2.02	10.75	0.251
σ	—	0.202	0.10	2.428	0.0658

^aDry feed.¹Closure was forced to 1 to determine (by difference) the H₂O output, which was not measured.²By difference.

Table 2. Continued

Run	Dry Air In (kg/kg DF)	Energy Yield (MJ /kg DF)	Total Energy Out (MJ /hr)	Mass Conversion Efficiency(%)
08/27	1.53	11.80	757.9	85.81
08/28	1.50	12.11	701.7	85.64
08/29	1.22	9.22	686.0	76.70
08/29	1.60	13.64	529.1	89.71
09/10	1.39	12.27	488.2	87.03
09/11	1.49	11.89	216.5	87.10
09/24	1.60	11.75	523.7	85.99
09/28	1.45	12.23	605.4	89.69
10/01	1.35	11.88	676.3	87.54
10/03	1.35	12.08	710.1	85.95
10/05	1.40	12.05	682.1	86.11
10/05	1.32	11.78	824.4	85.61
10/08	1.52	13.02	791.3	89.07
10/10	0.97	8.33	385.8	72.14
10/12	1.26	12.02	493.3	86.39
10/15	1.40	12.69	723.5	88.53
All data				
mean	1.40	11.80	—	85.56
σ	0.16	1.29	—	4.65
>95%				
mean	1.44	12.23	—	87.16
σ	0.10	0.54	—	1.51

Table 2. Continued

Run	Cold Gas Efficiency(%)	Highest Temp. Observed (°K)
08/27	69.49	1008.3
08/28	71.34	986.1
08/29	54.32	977.8
08/29	80.32	983.3
09/10	72.24	852.8
09/11	70.05	947.2
09/24	69.22	966.7
09/28	72.03	1172.2
10/01	69.97	1005.6
10/03	71.17	1005.6
10/05	70.95	994.4
10/05	69.37	983.3
10/08	76.69	1093.3
10/10	49.08	1025.0
10/12	70.78	1052.8
10/15	74.71	1166.7
All data		
mean	69.48	1013.9
σ	7.61	78.9
>95%		
mean	72.02	1015.6
σ	3.18	83.9

Table 3. Dry Gas Composition (%)

Run	H ₂	C ₂ H ₄	CO ₂	C ₂ H ₄	C ₂ H ₆	N ₂	CH ₄	CO
08/27	11.68	0.31	15.74	1.10	0.20	49.28	3.29	18.39
08/28	12.66	0.25	14.94	1.00	0.19	48.41	3.50	19.05
08/29	10.20	0.34	15.30	1.22	0.21	50.54	3.49	18.70
08/29	14.28	0.23	14.99	1.03	0.17	46.78	3.29	19.23
09/10	14.93	0.21	14.81	0.80	0.17	45.62	3.45	20.01
09/11	12.79	0.24	15.81	0.81	0.20	48.31	3.96	17.88
09/24	14.42	0.18	14.95	0.68	0.18	49.50	3.16	16.94
09/28	17.53	0.08	15.01	0.60	0.10	44.62	2.25	19.81
10/01	15.47	0.16	15.34	0.72	0.17	45.13	3.28	19.73
10/03	17.37	0.14	13.91	0.68	0.11	44.56	2.78	20.45
10/05	17.62	0.09	13.43	0.55	0.09	45.02	2.53	20.66
10/05	16.28	0.18	15.01	0.82	0.15	45.10	3.26	19.20
10/08	17.27	0.09	12.68	0.51	0.08	45.18	2.21	21.98
10/10	15.27	0.19	14.07	0.76	0.15	46.24	3.20	20.11
10/12	16.52	0.21	15.11	0.82	0.18	43.63	3.33	20.20
10/15	17.86	0.08	13.23	0.51	0.08	43.82	2.49	21.93
All data								
mean	15.13	0.19	14.64	0.79	0.15	46.36	3.09	19.64
σ	2.88	0.08	0.91	0.21	0.05	2.18	0.50	1.33
Chips								
mean	12.20	0.28	15.24	1.09	0.19	48.75	3.39	18.84
σ	1.71	0.05	0.37	0.10	0.02	1.58	0.11	0.37
pellets								
mean	16.11	0.15	14.45	0.69	0.14	45.56	2.99	19.91
σ	1.56	0.06	0.96	0.12	0.04	1.73	0.54	1.44

Table 4. Statistical Analysis

Feed	y	x	R ²	P>F-value	Significant Regression model
Chips	Char yield	Rotation of grate	0.99	0.0027	$y = 30.3(x-1.5)^{-0.911}$
Pellets	Char yield	Rotation of grate	0.72	0.0018	$y = 21.2(x-2.5)^{-0.304}$
Chips	Cold gas efficiency	Rotation of grate	0.98	0.0098	$y = 64.8(x-2.97)^{0.108}$
Pellets	Cold gas efficiency	Rotation of grate	0.91	0.0001	$y = 62.9(x-2.97)^{0.0679}$
Chips	Air input	Rotation of grate	0.998	0.0008	$y = 1.46(x-2.97)^{0.0506}$
Pellets	Air input	Rotation of grate	0.81	0.0004	$y = 1.19(x-2.9)^{0.0887}$
All	Cold gas efficiency	Char Yield	0.93	0.0001	$y = 86.3-1.37x$
All	Air input	Char Yield	0.79	0.0001	$y = 1.72-0.026x$
All*	Energy output	Dry feed rate	0.79	0.0001	$y = 18.6+11.9x$

*Excluding the two runs at the grate rotation of 3.0 min/rev.

Table 5. Comparisons of the Present Study with the Previous Study (Walawender et al., 1985).

Study		Present	Previous
Operating parameters:			
Feedstocks		Chips, pellets	Chips
moisture (wet %)		9.99 ,6.50	10.33
Bed support		W/B, WO1, WO	W/B
Grate rotation (min/r)		3-14	20-30**
Material balance closure (%)		94-106	80-115
Input: Dry feed rate (kg/hr)		18-74	27-126
Dry air in* (kg/kg DF)		1.44	1.60
Output: Gas yield* (m ³ /kg DF)		2.02	2.19
char yield* (kg/kg DF)		0.108	0.0334
water out* (kg/kg DF)		0.25	0.26
Mass conversion efficiency(%)		87.2	87.9
Energy aspects:			
Dry gas HHV (MJ/m ³)		6.08	5.51
Energy yield(MJ/kg DF)		12.2	12.0
Energy output (MJ/hr)		220-820	310-1430
Cold gas efficiency(%)		72.0	71.5
Gas composition: (%)			
Major components	H ₂	15.1	15.1
	CO	19.6	19.1
	CH ₄	3.09	2.51
	CO ₂	14.6	15.8
	N ₂	46.4	46.8

*Mean for the runs within of 5% closure.

**Estimated.

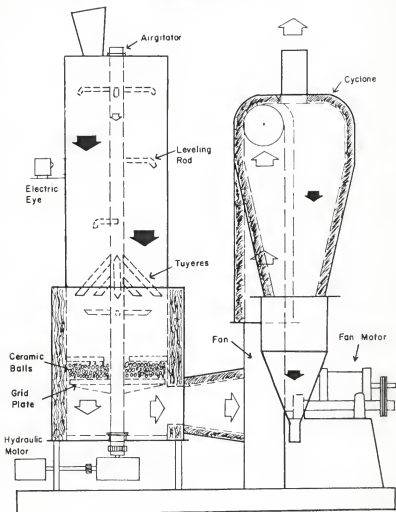


Figure 1. Commercial Gasifier.

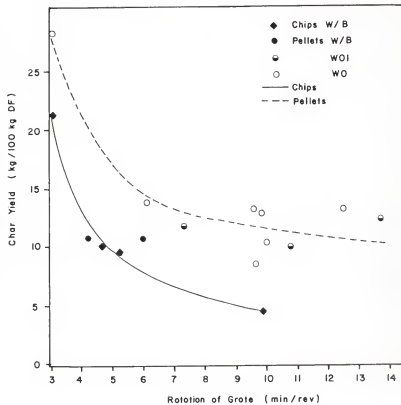


Figure 2. Char Yield versus Rotation of Grate.

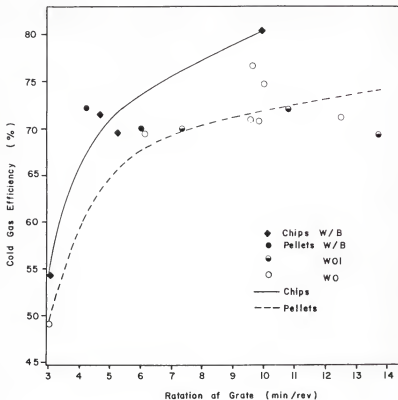


Figure 3. Cold gas Efficiency versus Rotation of Grate.

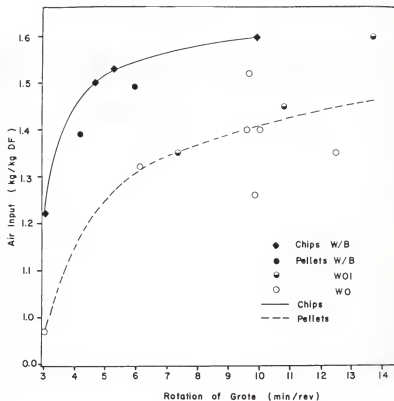


Figure 4. Air Input versus Rotation of Grotte.

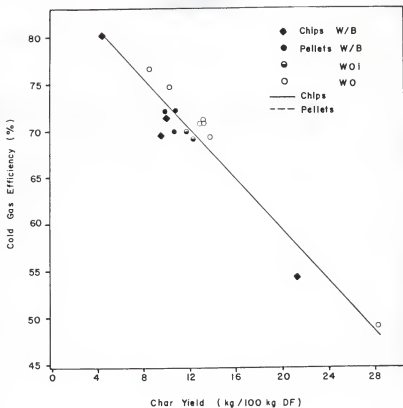


Figure 5. Cold gas Efficiency versus Char Yield.

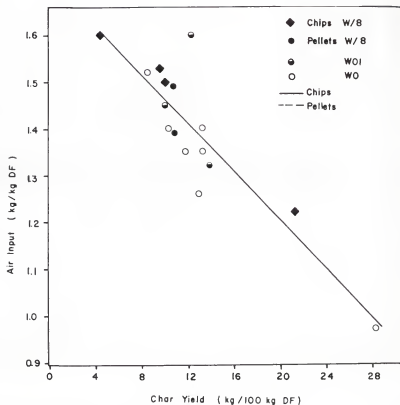


Figure 6. Air input versus Char Yield.

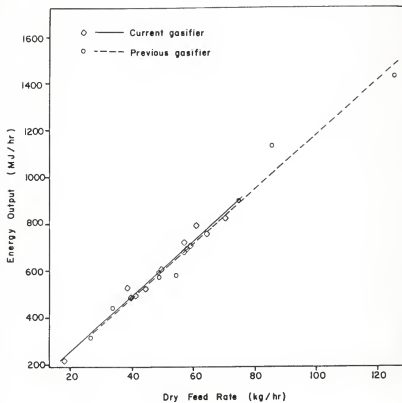


Figure 7. Energy Output versus Dry Feed Rate.

CHAPTER 6

CONCLUSIONS AND RECOMMENDATIONS

Air gasification of woody biomass was studied in two (0.6 m ID) commercial downdraft gasifiers of similar design. The objectives were to determine the gasifier performance and to investigate the influence of several operating variables. For the first gasifier, the grate rotational speed was proportional to the fan speed, with the fan speed being variable. For the second gasifier, the fan speed was fixed and the grate rotational speed was variable.

Studies with the first gasifier investigated the effect of feed rate with wood chips as the feedstock. The feed rate was proportional to the fan speed and was varied from 27 to 126 kg/hr. The second gasifier was used to investigate the effects of the following operating variables: 1) grate rotation speed, 2) type of feed material, and 3) bed support. The grate rotation speed was varied from 3 to 14 min/rev. and wood chips and wood pellets were used as feedstocks. Bed supports consisted of a 9 cm layer of ceramic balls (2 cm diameter) on top of the grid plate (grate), the grid plate with half of its available open area obstructed and the unobstructed grid plate.

Measured data included the feed rate, air input rate, gas output rate, gas composition, char rate, and condensate and tar rates. These data were used to evaluate material balance closures for the experiments. The data were also used to evaluate various gasifier performance measures, including the gas yield and gas heating value, the mass conversion and cold gas efficiencies, and the total energy output.

The major results from the experimental studies are outlined below.
For the first gasifier

1. Material balance closures on 16 runs ranged from 80% to 115%, with 9 runs within 5% of closure.

2. The gasifier performance was found to be remarkably consistent (in terms of various performance measures) over a four fold range of feed rate, with the energy output ranging from 320 to 1430 MJ/hr.

3. Various performance measures, when expressed on a unit mass of dry feed basis, were found to be independent of the feed rate. The mean for the gas yield was $2.19 \text{ m}^3/\text{kg}$; for the gas heating value, $5.51 \text{ MJ}/\text{m}^3$; for the char yield, it was $0.033 \text{ kg}/\text{kg}$; for the mass conversion efficiency, it was 87%; and for the cold gas efficiency, it was 72%.

4. The tar yield was negligible. The mean value for the tar yield was 0.0032 kg per kg of dry chips.

For the second gasifier

5. The char yield was significantly higher for pellets than for chips, due to the larger characteristic dimension for pellets relative to chips.

6. The char yield increased with increasing grate rotation speed, and the cold gas efficiency decreased with increasing grate rotation speed. This behavior was due to an inadequate air-to-feed ratio, a direct consequence of the fixed fan speed.

7. No significant effects were detected for variation of the type of bed support.

Eight data sets from the first gasifier (with material balance closures within 5%) were used to conduct a detailed mass and energy analysis. The mass analysis resulted in an empirical stoichiometry for the gasification of wood chips. This was accomplished through a two level adjustment of the original data which forced both the overall and elemental balances to perfect closure. The energy analysis was based on the thermodynamic first and second laws. It evaluated the first law and second law efficiencies for the process for various operation modes (combinations of different usable products and output temperatures). Various energy dissipations were also determined, including

the system heat loss, the exergy loss due to the heat loss and the exergy dissipation due to irreversibilities of the process.

The important results from the energy analysis are outlined below.

1. The highest first law and second law efficiencies (with all outputs as usable products at hot output) were 89.5% and 62%, respectively.
2. The lowest first law and second law efficiencies (dry cool gas) were 72% (agreed with the experimental observation) and 53%, respectively.
3. The system heat loss was 10.5% of the input energy (enthalpy), and the corresponding exergy loss was 7% of the input exergy.
4. The exergy dissipation due to various irreversibilities of the process was 31% of the input exergy.

The studies in this thesis have focussed on the material and energy balances of the process. Very little has been done to investigate transport phenomena in the system. (Measurement of the pressure drop across the bed was attempted; however, the data were not reported due to incomplete data.) The experimental study revealed that an inadequate air-to-feed ratio resulted in a decrease in the cold gas efficiency as the grate rotation speed increased. However, with the current gasifiers, it is impossible to control the air-to-feed ratio arbitrarily. Hence, it is desirable to modify the gasifier to have the air-to-feed ratio under control. In principle, the air input rate is determined by the pressure drop across the bed and the bed porosity. The bed porosity is in turn determined by the bulk density of the feedstock and the imposed agitation of the bed (e.g., by the rotation of the grate). The use of fan with powerful suction and variable speed will enable the control of the pressure drop across the bed. The use of feedstocks with various bulk densities and various degrees of bed agitation will enable the control of bed porosity. The control of the pressure drop and the bed porosity together

makes the control of air input rate possible and hence, the control of air-to-feed ratio.

Wood chips and wood pellets were gasified in this thesis. Other biomass materials, such as peat pellets and paper pellets, can also be used as feedstocks to investigate the gasifier performance over a wide variety of feed materials.

Although some temperature measurements were made at various positions in the system, the data were not sufficient to indicate the temperature distribution in the system. The measurement of the temperature distribution is desirable, because the temperature distribution indicates various active zones in the system. The measurements of product distribution and pressure distribution in the active zones are also useful to investigate the chemical processes and transport phenomena in the process, although these are more difficult to measure.

High char yield from the gasifier is undesirable. However, due to the nature of the downdraft gasifier, it is impossible to completely eliminate the char yield. The elimination of the char yield would be possible by introducing a secondary air (or oxygen) into the char bed. The char would be consumed partly by the oxidation reaction and partly by gasification reactions with the gaseous phase (using the heat released by the oxidation reaction). This concept originates from the updraft gasification which has no char yield.

THE GASIFICATION OF BIOMASS
IN COMMERCIAL DOWNDRAFT GASIFIERS

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Air gasification of wood chips and wood pellets was studied in two (0.6m ID) commercial downdraft gasifiers of similar design. The objectives were to determine the gasifier performance and the influence of several operating variables. For the first gasifier, the grate rotational speed was proportional to the fan speed, with the fan speed being variable. For the second gasifier, the fan speed was fixed and the grate rotational speed was variable.

Studies with the first gasifier investigated the effect of wood chip feed rate which was proportional to the fan speed. The chip rate ranged from 27 to 126 kg/hr. Measurements of the feed rate, gas composition, char rate, condensate rate, gas rate, and air rate were used to evaluate material balance closures. The data were also used to evaluate gasifier performance measures including the gas yield and the mass conversion and cold gas efficiencies. The performance measures, when expressed on a unit mass of dry feed basis, were found to be independent of the feed rate. The gas yield averaged 2.19 m³/kg and the char yield averaged 0.033 kg/kg while the mass conversion and cold gas efficiencies averaged 87% and 72% respectively.

Eight data sets were used to determine gasifier efficiencies based on the first and second laws of thermodynamics. As a part of the analysis, it was necessary to develop an empirical stoichiometry for the overall gasification. Primary and secondary adjustments were made on the data which forced both the overall and elemental balances to perfect closure. Efficiencies were determined for different classes of usable products. These included dry cool gas at one end and moist hot gas with both tar and char at the other. The first law efficiency (dry cool gas) was 72% and agreed with the experimental observation. The second law efficiency (dry cool gas) was 53% and the system heat loss was 10.5% of the input energy (enthalpy). The corresponding exergy

(available energy) loss was 7% of the input exergy and the exergy dissipation (due to irreversibilities) was 31% of the input exergy. With all of the products considered feasible, the first law efficiency was 89.5% and the second law efficiency was 62%.

The second gasifier was used to investigate the effects of the following operating variables: 1) grate rotation speed, 2) type of feed material, and 3) bed support. The grate rotation speed was varied from 3 to 14 min/rev. and wood chips and wood pellets were used as feedstocks. Bed supports consisted of a 9 cm layer of ceramic balls (2 cm diameter) on top of the grid plate (grate), the grid plate with half of its available open area obstructed and the unobstructed grid plate. No significant effects were detected for variation of the type of bed support. For both feed materials, it was observed that the char rate increased and the cold gas efficiency decreased with increasing grate rotation speed. This behavior was due to an inadequate air to feed ratio, a direct consequence of the fixed fan speed. The char yield was observed to be higher for pellets than for chips. This behavior was due to the larger characteristic dimension for pellets relative to chips.